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Fuel Quality/Processing Study

Volume I: Final Report

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Prepared for
NATIONAL AERONAUTICS AND SPACE ADMINISTRATION
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TABLE OF CONTENTS

ABBREVIATIONS

Atm, ATM	atmospheric
bbl	barrel(s)
BPD	barrels per day
BPSD	barrels per stream day
Btu	British thermal unit
cs, cst	centistokes
D	day
DCF	discounted cash flow
EDS	Exxon Donor Solvent (process)
EP	endpoint
FCC	fluid catalytic cracking (unit)
FCI	fixed capital investment
FOE	fuel oil equivalent
gal	gallons
gpm	gallons per minute
HHV	higher heating value
H.P.	high pressure
IBP	initial boiling point
kWh	Kilowatt hour(s)
LHV	lower heating value
LP	linear programming
L.P.	low pressure
LPG	liquefied petroleum gas
LTPD	long tons per day
M	thousand
max	maximum
min	minimum
mg	milligram(s)
MM	million
MMM	billion
N	Nitrogen
No.	number
ppm	parts per million
psi	pounds per square inch
psig	pounds per square inch gauge
RON	research octane number
RPSP	required product selling price
SCFD	standard cubic feet per day
SFV	Saybolt Furol Viscosity
SRC	solvent refined coal
ST	starting temperature
SUV	Saybolt Universal Viscosity
Temp	temperature
TBP	true boiling point
TPD	tons per day
Vac, VAC	vacuum
vol	volume
wt	weight
wt%	weight %

SECTION 1

INTRODUCTION

This report presents the results of Tasks I through V portion of the Fuel Quality/Processing Study project for production of gas turbine fuels. The objective of the study was to provide a data base to be used to establish an intelligent trade-off between advanced turbine technology and liquid fuel quality. Synthetic fuels (synfuels) to be emphasized include those derived from coal and shale.

The intent is to use the data base to be produced in this study to guide the development of specifications for future synthetic liquid fuels anticipated for use in the time period 1985-2000. It is also to be used as a basis for evaluating the value and benefits of federally sponsored R&D efforts in the field of advanced gas turbine technology.

The project assessed relative fuel costs, quality and energy efficiency for a number of fuel sources and processing alternatives. An objective was to accelerate implementation of fuel-flexible combustors for industrial and utility stationary gas turbine systems. This is to be accomplished in the broader U.S. Department of Energy (DOE) Low NOx Heavy Fuel Combustor Program by generating and demonstrating the technology base for development of reliable gas turbine combustors which are capable of sustained environmentally acceptable operation when using minimally processed synthetic fuels.

Work on this program was done for NASA-Lewis Research Center under contract DEN3-183. NASA's guidance in the performance of this study was most helpful and we express our appreciation.

The program structure consisted of five technical performance tasks which are briefly defined as:

TASK I - LITERATURE SURVEY

Define the properties and characteristics of near-future (1985-2000 time period) petroleum and synfuels, synfuels processes using coal or oil shale, fuel additives, on-site treatment processes and exhaust gas clean-up processes.

TASK II - ON-SITE PRETREATING

Evaluation of fuel treatment requirements and relative costs of pretreating and processing requirements for various levels of fuel impurity removal and fuel throughput.

TASK III- EXISTING REFINERIES TO UPGRADE FUELS

Investigation of feasibility and relative costs of upgrading oil shale derived and coal direct liquefaction synfuels in existing refinery complexes.

TASK IV - NEW REFINERIES TO UPGRADE FUELS

Definition of the technical capability and economics of new refinery processes and/or refineries and/or integrated "confiners" to produce acceptable gas turbine fuels from oil shale derived and coal direct liquefaction synfuels.

TASK V - DATA EVALUATION

Evaluation of results obtained from program Tasks I, II, III, and IV.

The Task I Literature Survey was transmitted to NASA-Lewis Research Center in April, 1980; it is presented here as an APPENDIX to this final report as a separate volume. The results from the Tasks II through V program are presented in the report sections which follow.

SECTION 2

SUMMARY

This section summarizes the key program results for the following subject areas:

- Literature Survey
- On-Site Fuel Pretreatment
- Existing Refineries to Upgrade Fuels
- New Refineries to Upgrade Fuels
- Environmental Considerations

An inhouse linear programming model served as the basis for determining economic processing paths for the existing refineries and new refineries syncrude upgrading. This involved development of extensive input data comprised of fuel properties, yields, component blending characteristics, incremental capital and operating costs, feed and product costs.

Economics are based on March, 1980 price levels. This applies to estimated fixed capital investments (FCI), operating costs and required product selling prices (RPSP). RPSPs are based on a 15% discounted cash flow (DCF) rate of return for all operations.

2.1 LITERATURE SURVEY

The volume entitled Task I - Literature Survey completed in April, 1980 is presented as the Appendix to this Fuel Quality/Processing Study report. Much of the information contained in this survey summary was used as reference material for completion of the subsequent tasks of this Fuel Quality/ Processing Study.

2.2 ON-SITE FUEL PRETREATING

Section 3 of this report summarizes the results of three process procedures consisting of water wash systems for reduction of alkali metals content of gas turbine liquid fuels prior to use. These are:

(1) Conventional wash system using electrostatic precipitation

(2) Possible alternate continuous centrifugal contactors

(3) Expansion of conventional wash systems by addition of continuous centrifugal contactors

The procedures, facilities, and economics for three (3) separate levels of alkali metal contamination were assessed; these were 20 ppm max, 20-200 ppm, and 200-2000 ppm. Each of these systems also include heating and filtration equipment for achieving operable fuel viscosity levels and removal of particulates, respectively.

The use of NOx removal processes to achieve permissible gas turbine/ waste heat boiler stack effluents was investigated. Development status for the large gas turbine effluent volumes is unfavorable. Accordingly, we suggest that fuel bound nitrogen content be reduced, along with aromatics and cyclic compounds, in the refinery upgrading processes.

Suppliers of conventional and continuous centrifugal contactor water wash equipment and systems were given copies of fifty-eight data sheets, from the Literature Survey Appendix report, representing a variety of synfuel liquids considered to be gas turbine fuel candidates. The assessments made regarding suitability of their equipment for processing the resid, oil shale and coal derived liquid fuels were:

- Conventional Systems - 40% of the fuels would present problems.
- Centrifugal Contactors - 16% of the fuels would present problems which could probably be circumvented.

Preliminary assessment of the costs of water washing indicate that they are in the range of 20-30 cents per barrel, exclusive of treatment for vanadium. Cost estimates and analyses indicate fixed capital investment and operating cost for the process portions could possibly be approximately 30% lower for the alternate continuous centrifugal contactor systems than for the conventional wash system using electrostatic precipitation.

2.3 EXISTING REFINERIES TO UPGRADE FUELS

Section 4 of this report describes a basic 200,000 barrel per day (BPD) representative petroleum refinery. An operation is defined in which the necessary equipment required to process the various oil shale and coal derived individual raw syncrudes compatibly at the rate of 50,000 BPD is added to the existing refinery while petroleum feed is reduced to maintain a normal product slate. Linear programming models were developed to quantify the description of the facilities and operation. Input files were prepared which comprised capital and operating cost items, fuel characteristics, raw material feed costs, utilities, product slates and their market values. The linear programming computer runs determined the optimum economic process path. Hydrotreating is a necessary processing step common to upgrading of alternate syncrudes. Reaction with hydrogen serves to reduce fuel bound nitrogen, sulfur and aromatics contents and to increase fuel stability.

A minimum of seven scenarios, with a total of 20 case and turbine fuel product variations, were developed for this phase of the study. These can be summarized as follows:

(1) Existing Refinery Normal Operation

(2) Shale Oil Upgrading
  a. Hydrotreating raw feed before distillation and subsequent processing.
  b. Distillation of raw feed before hydrotreating and other processing of distillation products.

(3) H-Coal Liquid Upgrading

a. Hydrotreating raw feed before distillation and subsequent processing.

b. Distillation of raw feed before hydrotreating and other processing of distillation products.

(4) SRC-II Liquid Upgrading

a. Hydrotreating of the total 950°F minus portion of the SRC-II liquid feed before distillation and subsequent processing.

b. Distillation of the total SRC-II liquid into cuts before hydrotreating and other processing.

Computer input diagrams are presented in Section 4, depicting the processing of each of the syncrude feeds.

The Exxon Donor Solvent (EDS) syncrude was not assessed due to limitation of time and cost resources. We judge that the properties of EDS would fall between those of H-Coal and SRC-II syncrudes, and therefore the results of an EDS assessment would be expected to fall between those of H-Coal and SRC-II syncrudes.

2.4 NEW REFINERIES TO UPGRADE FUELS

Suggested configurations were developed for new grass roots "stand alone" refineries processing 50,000 BPD of syncrude without petroleum crude feedstock. In these cases, a sizeable hydrogen facility must also be added. Since the severity of hydrotreating and hydrocracking the syncrudes is necessarily greater than is required by the refinery operating on petroleum crude feed, capital costs and operating costs are proportionally higher. The smaller capacity, 50,000 versus 200,000 BPD, constitutes a further proportionally higher cost for the "stand alone" syncrude refinery.

Six major scenarios, with a total of 15 cases and turbine fuel product variations, are involved in the upgrading of shale oil, H-Coal and SRC-II liquids in the new "stand alone" refineries, similar to the scenarios outlined under subsection 2.3.

Table 2-6 summarizes the parameters used for the 36 scenarios developed for this study. Variables included type of feedstock, refinery configuration (modified existing and new "stand-alone" refineries), hydrotreating before or after distillation, and turbine fuel specifications. Key characteristics of the turbine fuel specifications used are summarized in Table 2-7.

Comparative technical and economic information is contained in the Data Evaluation section, Section 6 of this report. Table 2-1 at the end of this section summarizes fixed capital investments for pro[illegible] units and equipment added to the existing refinery and equipment for new syncrude refineries. Data for Table 2-1 are contained in Tables 6-3, 6-9, 6-11, 6-16, 6-17, 6-22, 6-28, 6-29, and 6-34 for the three syncrude feeds: shale oil, H-Coal and SRC-II oils, for Case 1 and Case 2 modes of operation. Case 1 refers to operation in which the raw syncrude feed to the refinery is hydrotreated before distillation. Case 2 hydrotreats after distillation.

The capital costs of the "stand alone" syncrude refineries to process 50,000 BPD of syncrude are indicated to be of the order of twice that required for equipment added to the existing refinery to process the same amount of syncrude.

An exception is found for SRC-II "stand alone" refinery in Table 2-1, Case 2, T12. Here the fixed capital investment is the same for the existing and new refineries at about 75 million dollars. This change in equipment cost results from allowing a higher boiling point distillate for turbine fuel T12 in the new SRC-II refinery. Directionally, this specification change results in lower priced turbine fuel which indicates the impact of deleting refinery units.

Table 2-2 is a summary of RPSP from turbine fuel, expressed as dollars per barrel, produced in the syncrude plus petroleum crude feed refineries and the "stand alone" syncrude refineries. The data are from the required revenue summaries in Tables 6-4, 6-10, 6-14, 6-15, 6-23, 6-26, 6-27 and 6-35. The required revenue shown is the selling price per barrel of turbine fuel required to maintain the refineries' normal profitability of 15% discounted cash flow at no penalty to other products. The required revenue is based on raw syncrude costs to the refinery which were selected from published information; it is nevertheless arbitrary. The sensitivity of RPSP to syncrude prices was later developed.

The saleable products normally produced in the refineries are:

(1) Non-Leaded Gasoline
(2) No. 2 Fuel Oil
(3) No. 6 Fuel Oil
(4) LPGs
(5) Coke
(6) Byproduct Sulfur and Ammonia

The assessment envisioned these products to be sold at published market prices. The estimated gas turbine fuels' high required unit selling prices result from a combination of factors:

- Syncrude feed price is high.
- Severity of hydrogen treatments exceeds that for petroleum crudes necessitating greater quantities of hydrogen.
- Operations are capital intensive requiring more costly equipment than required for average petroleum operations.

Turbine fuels produced from syncrudes in an existing refinery are estimated to have a required revenue ranging from \$29 to \$44 per barrel for the parameters used for this study. This compares with the \$23 and \$32 per barrel market price for No. 6 fuel oil and No. 2 oil, respectively. Required revenue for the "stand alone" syncrude refineries range from \$67 to \$155 per barrel, which is definitely beyond current market prices. This indicates refining costs for processing syncrudes while producing conventional products are higher than for petroleum refining, in all cases.

2.5 DATA EVALUATION

Review of the linear programming results, as summarized in Tables 2-1 and 2-2, indicates that upgrading of syncrudes might be done at lower cost in existing large petroleum crude refineries rather than in new refineries designed for synthetic crude processing. Incremental capital investment for the former is lower - about half of that required to install a new 50,000-BPD refinery to process the same quantity of synthetic crude.

The operating costs for processing the 50,000 BPD of synthetic crude through the 200,000-BPD petroleum refinery along with petroleum are considerably lower than for the alternative syncrude refinery. Required product revenues for gas turbine fuels is approximately one-third that required for a new synthetic crude refinery, based on use of this study's parameters and procedures.

The study indicates the highest capital investment addition to the existing refinery is required for processing shale oil, the lowest for SRC-II and that for H-Coal processing in between. The lowest overall operating costs are achieved by the shale oil cases with H-Coal and SRC-II operating costs being comparable. These are related to the feed costs used. The appreciably lower shale oil feed cost differential more than compensates for the higher FCI addition. The comparison is as follows:

Feed	Feed Cost Used ($ per barrel)	FCI Average ($ million)	Rounded Average Turbine Fuel Required Revenue ($ per barrel)
Shale Oil	25	215	32
H-Coal Oil	32	121	40
SRC-II Oil	30	95	41
Petroleum Crude	30	--	--

The fixed charge for petroleum crude is shown to indicate its relationship to the synthetic crude feed costs.

Operation of the existing refinery on the combination of petroleum crude and synthetic crude results in a reduction of the normal 200,000 BPD petroleum crude feed by 30,000 to 40,000 BPD while maintaining the near normal gasoline and other main products output plus the production of 20,000 BPD of gas turbine fuel.

The study results indicate that the processing of syncrudes in an existing large petroleum refinery, with the addition of equipment as required for the synfuels processing, is the most economical route. Processing through a new smaller syncrude refinery is more costly. A comparison summary of these factors and feed costs is as follows:

	Feed Cost ($/bbl)	FCI Average ($ Million)		Turbine Fuel Required Revenue ($ per barrel)	
		Existing[a] Refinery	New[b] Refinery	Existing Refinery	New Refinery
Shale Oil	25	215	488	32	103
H-Coal Oil	32	121	247	40	101
SRC-II Oil	30	95	213	41	119
Petroleum Crude	30	--	--	--	--

[a] FCI of process unit additions to a 200,000 BPD petroleum refinery having a base FCI of approximately $600 million.

[b] FCI of process units for refining 50,000 BPD of syncrudes.

Sensitivities were developed for RPSP to (1) raw synfuel cost to the refinery, and (2) fixed capital investment for the refineries. Availability of the sensitivity values provides the reader flexibility to determine the effect of differing syncrude values and facilities costs on synthetic turbine fuel values. Results showing the sensitivities are presented in tabular form at the end of this Summary section.

The sensitivity assessment results indicate required product selling price (RPSP) to be more sensitive to syncrude feed cost than to total capital investment costs. Roughly, for the "stand alone" new refinery, a change of $1

per barrel syncrude cost results in a turbine fuel required selling price change of $8-10 per barrel. For existing refineries, RPSP is changed about $2.50 per barrel per $1 change in syncrude cost.

2.6 ENVIRONMENTAL CONSIDERATIONS

Section 7 presents an outline of current emission standards. The upgrading hydrotreating processing serves to reduce sulfur and fuel bound nitrogen content of the gas turbine fuels, and other process streams, so that fuel maximum sulfur and nitrogen contents of 0.8% and 0.25%, respectively, can be met.

The upgrading of the synthetic crudes through hydrotreating reduces their polycyclic and aromatic hydrocarbon content. This represents a reduction of contained carcinogens, thus reducing the biohazards of the syncrude based intermediates and products.

Table 2-1 - Fixed Capital Investment
Onsite Facilities, $ Million

Synfuel	Process Units Added To Existing Refinery					Process Units for New Synfuel Refinery					
	TF1	TF2	TF3	T11	T13*	TF1	TF2	TF3	T11	T12	T13
Case 1:											
Shale Oil	237	--	236	--	--	500	--	--	--	--	--
H-Coal	152	152	152	--	--	267	236	--	--	--	--
SRC-II	139	--	130	--	126	263	270	277	--	--	278
Case 2:											
Shale Oil	236	--	209	189	184	497	--	479	486	--	476
H-Coal	107	--	113	82	87	239	--	--	--	--	--
SRC-II	55	58	84	--	75	245	112	--	--	75	--

Table 2-2 - Turbine Fuel Required Revenue
$ per Barrel

Synfuel	Existing Refinery Plus Synfuel Feed					New Synfuel Refinery					
	TF1	TF2	TF3	T11	T13	TF1	TF2	TF3	T11	T12	T13
Case 1:											
Shale Oil	33	--	33	--	--	116	--	--	--	--	--
H-Coal	44	44	43	--	--	121	114	--	--	--	--
SRC-II	45	--	42	--	40	151	151	150	--	--	150
Case 2:											
Shale Oil	34	--	32	30	29	103	--	98	101	--	98
H-Coal	39	--	39	37	36	67	--	--	--	--	--
SRC-II	42	40	39	--	37	155	119	--	--	107	--

* See Table 6-1, page 6-14, for turbine fuel specifications. The specifications differ in such characteristics as nitrogen content, boiling point range and viscosity.

Table 2-3 - Required Product Selling Price Sensitivities
Case 2, Turbine Fuel TF1
$ per barrel

Syncrude Feed	Existing Refinery -30%	Existing Refinery Base Case	Existing Refinery +30%	New Syncrude Refinery -30%	New Syncrude Refinery Base Case	New Syncrude Refinery +30%
Sensitivity to Total Capital Investment:						
Shale Oil	29	34	40	57	103	149
H-Coal	37	39	41	49	67	84
SRC-II	41	42	43	132	155	178
Sensitivity to Syncrude Feed Cost:						
Shale Oil	15	34	54	26	103	180
H-Coal	15	39	63	- 8	67	141
SRC-II	20	42	65	65	155	245

Table 2-4 - Sensitivity Ratios of Turbine Fuel Required Product Selling Price (RPSP) to Fixed Capital Investment (FCI)

Syncrude	Existing Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% FCI}}$	New Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% FCI}}$
Shale Oil	0.183	1.53
H-Coal	0.067	0.58
SRC-II	0.033	0.77

Table 2-5 - Sensitivity Ratios of Turbine Fuel Required Product Selling Price (RPSP) to Syncrude Feed Cost

Syncrude	Existing Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% Syncrude Cost}}$	New Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% Syncrude Cost}}$
Shale Oil	0.65	2.57
H-Coal	0.80	2.48
SRC-II	0.75	3.00

Table 2-6 - Synthetic Turbine Fuels Production/Refining Scenarios Analyzed

Refinery Feedstock	Refinery Configuration: Existing	Refinery Configuration: Modified Existing	Refinery Configuration: New "Stand Alone"	Operations Mode: Case 1 Hydrotreat Whole Feed Before Distillation	Operations Mode: Case 2 Hydrotreat Individual Cuts After Distillation	Product Specifications
Petroleum Crude	1					1
H-Coal Liquids + PC		7		2	4	5
H-Coal Liquids			3	2	1	2
SRC II Liquids + PC		7		3	4	4
SRC II Liquids			7	4	3	4
Shale Oil + PC		6		2	4	4
Shale Oil			5	1	4	4

Total Scenarios Analyzed = 36

Table 2-7 - Synthetic Turbine Fuel Specifications

Specification Designation	Distillation End Point	Sulfur, % Max	Nitrogen, % Max	Viscosity @ 100°F Min, CST	Viscosity @ 100°F Max, CST
TF1	650°F	0.7	0.25	1.8	5.8
T11	650°F	0.7	1.0	1.8	5.8
TF2	<1000°F	0.7	0.25	1.8	30.0
T12	<1000°F	0.7	1.00	1.8	30.0
TF3	>1000°F	0.7	0.25	1.8	160
T13	>1000°F	0.7	1.0	1.8	160
TF4	>1000°F	0.7	0.25	1.8	900

SECTION 3

ON-SITE FUEL PRETREATING

The gas turbine is a high speed, high temperature machine whose life and performance is vulnerable to foreign fuel constituents, even in trace quantities. Accordingly, on-site pretreatment of even the best grades of fuel is common utility and industrial practice. The advent of coal and shale derived liquids and even petroleum resids introduce the possibility of the presence of new and increased quantities of harmful impurities. It is the purpose of this section to present the methods of conventional pretreatment and discuss the possible new requirements which may be introduced by new sources of gas turbine fuels.

3.1 GAS TURBINE FUEL IMPURITIES

The impurities considered objectionable and which are limited in quantities by accepted specifications[1,2,3,4] are summarized and described below:

- <u>Particulates</u>. Combustible and non-combustible material which is suspended in the fuel which can cause deposition on turbine blades and can contribute to blade corrosion and/or erosion.

- <u>Alkali metals</u>. Sodium and potassium combine with vanadium to form low melting salts which are corrosive to the turbine blades. Calcium causes hard-bonded deposits on the turbine blades which are difficult to remove.

- <u>Vanadium</u>. Forms molten vanadium pentoxide which causes severe corrosion of gas turbine blades.

- <u>Lead</u>. Causes corrosive deposits and also inhibits beneficial effects of vanadium anti-corrosion additives. However, lead is not expected to be present in synfuels.

- **Copper**. An oxidation catalyst causing poor fuel thermal stability. Copper is not expected to be present in synfuels.

- **Sulfur**. On combustion contributes to objectionable SO_2 emissions.

- **Nitrogen**. Fuel bound nitrogen contributes to nitrogen oxide pollutants in exhaust gases, adding to those formed from nitrogen in air during combustion.

Of the above, all but the last two impurities are usually rendered unobjectionable by on-site fuel pretreatment. It must be noted that up to 4% sulfur does not affect performance or have an adverse effect on the gas turbine components.

Use of NOx removal processes on gas turbine/waste heat boiler emissions was investigated. Degree of development is limited and pertains to conventional steam boiler rather than gas turbine operation. The major drawback for gas turbine application is the extremely large and expensive catalyst chamber due to the large gas volume per kilowatt generated. Accordingly, it was considered fuel bound nitrogen could be reduced in the refinery upgrading process.

Turbine manufacturers have formulated fuel specifications pertinent to optimum operation and compatible with their machines. Tables 3-1, 3-2 and 3-3 are tabulations of liquid fuels specifications of three major U.S. gas turbine manufacturers. Accordingly, specification items not met by the delivered turbine fuels are corrected by appropriate pretreatment procedures.

3.2 CONVENTIONAL PRETREATMENT METHODS

The following summarize conventional pretreatment methods currently in use for systems burning petroleum based gas turbine fuels.

Figure 3-1 is a simplified diagram of a conventional two-stage electro-

static precipitator fuel pretreatment system. This system will satisfactorily handle fuels having less than 200 ppm alkali metals content.

3.2.1 ASH AND PARTICULATE REMOVAL

Minor quantities of ash and particulates such as scale particles from tanks and piping are removed by filtration. Filters are standard equipment in the fuel feed circuit to all gas turbines and should be capable of removing material down to at least ten microns in size.

3.2.2 VISCOSITY CORRECTION

Viscosity can usually be corrected, as necessary, by heating. Heaters are standard equipment included in the pretreatment systems.

3.2.3 ALKALI METALS REMOVAL

These impurities are removed by water washing, using high quality clean water. Many gas turbines are provided with heat recovery steam generators. In these cases wash water is provided from the boiler feedwater make-up system, usually evaporated or deionized. The use of normal potable water may be unacceptable because sodium salt content could further contaminate the fuel.

The wash water and oil are contacted in stationary line mixers such as eductors, mixing Ts or valves. Wetting agent is injected ahead of the mixing point for easier contacting of the fuel oil and wash water. The mixture is then passed through low velocity treater tanks in which the separation of oil and water is effected by electrostatic precipitation. This separation is sometimes accomplished using centrifuges alone, and also in combination with electrostatic precipitators.

Figure 3-2 is a plot plan of the typical on-site fuel pretreating system portrayed in Figure 3-1.

3.2.4 VANADIUM INHIBITING

The detrimental effects of vanadium in excess of 2 ppm in the fuel can be inhibited by the addition of magnesium compound solutions formulated using various vehicles. Generally 3 ppm of inhibitor solution is utilized per 1 ppm of contained vanadium.

3.3 PRETREATMENT COSTS

The costs involved in the pretreatment of gas turbine fuels will vary in accordance with the level of impurities contents. Three levels of alkali metals content are considered for this study with an oi processing rate of 300 gallons per minute or approximately 10,000 BPD, equivalent to 3 million barrels per year based on an 80% equipment load factor. This would supply fuel to a nominal 200 megawatt power generating plant operating at base load.

Concentration levels included are:

- To 20 ppm
- Over 20 to 200 ppm
- Over 200 to 2000 ppm

Each of these levels requires a different size system for proper reduction of contained alkali metals to a maximum of 3 ppm of combined sodium, potassium and calcium. Fixed capital investment and operating costs will accordingly vary.

3.3.1 FIXED CAPITAL INVESTMENT

The estimated fixed capital investments for required conventional pretreatment process systems are summarized below with investments expressed in March, 1980 dollars, exclusive of laboratory and fuel supply storage tanks.

Alkali Metals (ppm)	System Required	Fixed Capital Investment ($ Thousand)
To 20	2 stage	1,680
Over 20 to 200	3 stage	2,050
Over 200 to 2000	4 stage	2,560

The two, three and four stage systems would be adequate for higher levels of alkali than shown, namely 150, 500 and 2000 ppm for 29°API fuels. However, practice has been to provide additional capacity as a precautionary measure against varying fuel deliveries.

3.3.2 Operating Costs

The estimated annual operating costs and the average cost per barrel (treated at an 80% load factor) for the above systems are:

Alkali Metals (ppm)	Direct ($Thousand)	Indirect ($Thousand)	Annual ($Thousand)	Average per barrel cost (cents/bbl)
To 20	286	302	588	19.6
Over 20 to 200	320	369	689	22.9
Over 200 to 2000	358	461	819	27.3

The details for these estimates are presented in Table 3-4. Electricity, steam and wetting agent are principal utilities and material costs for the water wash pretreatment. The computations are based on an 80% load factor. Vanadium inhibitor costs are shown separately on a unit basis and must be added to the annual cost shown above to obtain the total.

3.4 SYNCRUDE PRETREATMENT ASSESSMENT

We conferred with manufacturers of fuel treating equipment and systems concerning performance of their equipment relative to use of coal and shale derived liquids as gas turbine fuels. Since such experience is lacking we submitted a total of fifty-eight coal and shale derived liquids and resid potential gas turbine fuel property data sheets, developed as part of Task I, asking for their opinion regarding suitability of their equipment and systems for water washing these fuels. Copies of these sheets are included as tables in the Literature Survey Appendix to this report. They are located in Section 2, Section 3 and Appendix B of the Literature Survey. The fifty-eight data sheets are identified in Table 3-5 at the end of this section.

3.4.1 CONVENTIONAL EQUIPMENT AND SYSTEMS

The following tabulation summarizes the opinion of a prominent manufacturer regarding the "suitability for washing" using electrostatic precipitator equipment with reference to the fifty-eight fuels property sheets submitted:

Opinion	Type	Number	% of Total
No Difficulties	Distillates & Blends	35	60
Possible Problems	Coal and Shale Oil Heavy Distillates and Petroleum Resid	3	5
Unable to Process	Heavy Coal, Shale Oil Fractions and Petroleum Resid	20	35
Total		58	100

The major hindrance to processing was fuel specific gravities being nearly equal to or greater than that of water. Satisfactory separation and water removal may be difficult in the equipment normally used.

In general, most of the coal and oil shale derived fuels could be handled in the conventional pretreatment equipment and systems. Accordingly, the cost information presented in subsections 3.3.1 and 3.3.2 applies to the fuels in the "No Difficulties" category. However, with 40% of the fuels assessed as presenting problems in water washing in conventional equipment, there was cause for concern. Alternate equipment, more adaptable to the new fuels, was deemed to be desirable.

3.4.2 ALTERNATE EQUIPMENT AND SYSTEMS

We considered that a continuous centrifugal extractor could be advantageously used for water washing pretreatment for reduction of alkali metals. This machine is widely used for similar extraction operations. The petroleum industry uses these for the solvent extraction step in the manufacture of lubricating oils.

The expectation is that the centrifugal extractor with its multi-stage contacting and separation feature might more effectively perform the water washing functions, particularly with respect to the heavy distillates, heavy fractions and resids.

Discussions with a centrifugal extractor manufacturer were held. To their knowledge, none of their machines are in turbine fuel water washing service. They were of the opinion that their machine would be applicable. They are in the process of exploring this application and are desirous of running tests in their pilot units. It was deemed advisable to explore this application in view of the opinion that nearly forty percent of the likely list of possible future gas turbine fuels might not be amenable to satisfactory processing in the conventional electrostatic precipitator systems.

The fifty-eight syncrude derived and resid fuel property data sheets, listed in Table 3-5, were also sent to the manufacturer of the centrifugal contactor. Their assessment of the applicability of their machine to the fifty-eight fuels is summarized as follows:

Opinion	Type	Number	% of Total
No Difficulties	Sp. Gr. differences 0.02 or greater	49	84
Requires slight dilution with lighter oil	Sp. Gr. differences less than 0.02	9	16
Total		58	100

Since the consensus is that there is a possibility that centrifugal contacting and separation may be advantageously used for water washing new fuels, preliminary economics were derived. These are considered only indicative of the possibilities, requiring confirmation by subsequent test work.

A single contactor, because of the multi-stage operating effect, will properly handle fuels containing up to 200 ppm alkali metals. Two machines would be required for levels to 2000 ppm. Diagrams of these two systems are shown as Figures 3-3 and 3-4 respectively.

Estimated fixed capital investment (FCI) costs for these systems compared to conventional installations are summarized below:

			FCI	
Alkali Metals (ppm)	Conventional System	Alternate System	Conventional System $ Thousand	Alternate System $ Thousand
To 20	2 stage	1 Contactor	1,680	1,200
20 to 200	3 stage	1 Contactor	2,050	1,200
200 to 2000	4 stage	2 Contactors	2,560	1,725

Estimated annual operating costs for the alternate centrifugal contactor systems, compared to conventional installations also relate favorably:

Alkali Metals (ppm)	Operating Costs			
	Total Annual		Average/bbl	
	Conventional ($ Thousand)	Alternate ($Thousand)	Conventional (cents/bbl)	Alternate (cents/bbl)
To 20	588	432	19.6	14.4
20 to 200	689	432	22.9	14.4
200 to 2000	819	563	27.3	18.7

The use of wetting agent should not be necessary for the centrifugal contactor system. This amounts to 0.7 to 0.9 cents per barrel and is included in the conventional system operation only. Details for the alternate centrifugal contactor case estimates are presented in Table 3-6.

The above figures indicate the alternate centrifugal contactor system to be worthy of further in-depth investigation.

The advent of coal, shale and resid-derived turbine fuels would, in some cases, require expansion of existing conventional systems. This might be accomplished by the installation and operation of a centrifugal contactor in conjunction with the existing conventional system. This concept is shown in Figure 3-6, wherein a two-stage system is augmented by a single centrifugal contactor unit in order to increase capacity from 200 ppm alkali metals content fuel to fuel containing 2000 ppm.

Equivalent total fixed capital investment and operating cost for this combination system, operating at 300 gpm and 80% load factor, would be expected to result in net overall lower costs compared to expansion by addition of two conventional stages:

Alkali Metals (ppm)	Fixed Capital Investment	
	Conventional 4-stage ($ Thousand)	Combination 2-stage plus Contactor ($ Thousand)
2000	2650	2335

Alkali Metals (ppm)	Operating Costs: Total Annual: Conventional 4-stage ($ Thousand)	Operating Costs: Total Annual: Combination 2-stage plus Contactor ($ Thousand)	Operating Costs: Average/bbl: Conventional (cents/bbl)	Operating Costs: Average/bbl: Combination (cents/bbl)
2000	819	748	27.3	24.9

3.5 SECTION 3 LITERATURE CITED

1. ANSI/ASTM D2880-78, Standard Specification for GAS TURBINE FUEL OILS.

2. Gas Turbine Liquid Fuel Specifications, GEI-41047G, General Electric Company. (Latest specification obtained 1980).

3. Westinghouse Liquid Fuels Specification, Westinghouse document 971021. (Received 1979).

4. Pratt and Whitney Aircraft GTF Specifications.

Table 3-1 - General Electric Liquid Fuel Specifications For Gas Turbines

Property	ASTM Test Method	Distillates Light	Distillates Heavy	Crudes and Blended Residual Fuels	Heavier Residual Fuels
Specific Gravity, 60°F	D1298	Report	Report	0.96	0.96
Kinetic Viscosity, cs, 100°F, min	D445	0.5	1.8	1.8	1.8
Kinetic Viscosity, cs, 100°F, max	D445	5.8	30	160	900
Kinetic Viscosity, cs, 210°F, max	D445	--	4	13	30
Flash Point, °F, min	D93	Report	Report	Report	Report
Distillation Temp, 90% Point, °F, max	D86	650	Report	--	--
Pour Point, °F, max	D97	0	Report	Report	Report
Carbon Residual (10% Bottoms), Wt %, max	D524	0.25	--	--	--
Carbon Residual (100% sample), wt %, max	D524	1.0	1.0	1.0	--
Ash, ppm, max	D482	50	50	Report	Report
Trace Metals, ppm, max					
Sodium Plus Potassium	--	1	1	1	1
Lead	--	1	1	1	1
Vanadium (untreated)	--	0.5	0.5	0.5	0.5
Vanadium (treated 3/1 wt ratio mg/Vol)	--	--	--	100	500
Calcium	--	2	2	10	10
Filterable Dirt, mg/100 ml, max	D2276	4	40	Report	Report
Water & Sediment, Vol %, max	D1796	0.1	0.1	1.0	1.0
Thermal Stability, Tube No., max	D1661	--	2	2	2
Fuel Compatibility, Tube No., max (50/50 mix with second fuel)	D1661	--	2	2	2
Sulfur, wt %, max[a]	D129	0.5	0.5	1.0	1.0
Hydrogen, wt %, min	--	12.0	12.0	11.3	11.3
Nitrogen, wt %, max	--	Fuel-bound nitrogen may be limited to meet any applicable codes on total NO_x emission.			

[a] Or compliance to any applicable codes.

Table 3-2 - Westinghouse Fuel Specification

Property	Distillate Fuel	Residual Fuel
Gravity, °API	26 min	12 min
Viscosity		
SUV at 100°F	32-45	-
SFV at 122°F	-	300 max
SUV at 210°F	-	220 max
Distillation, °F		
90% Evaporation	675 max	-
Water and Sediment, wt %	-	1.0 max
Ash, wt %	0.01 max	0.1 max
Metals - No Treatment		
Sulfur, wt %	2.0 max	4.0 max
Vanadium, ppm	2.0 max	5.0 max
Sodium, ppm	2.0 max	5.0 max
Calcium, ppm	10.0 max	10.0 max
Metals - Additive treatment for Vanadium Content		
Sulfur, wt %	-	4.0 max
Vanadium, ppm	-	200 max
Sodium, ppm	-	10 max
Calcium, ppm	-	10 max
Metals - Treatment required for Both Vanadium and Sodium		
Sulfur, wt %	-	4.0 max
Vanadium, ppm	-	200 max
Sodium, ppm	-	60 max[a]
Calcium, ppm	-	10 max

[a] Sodium must be reduced to 10 ppm max by water washing.

Table 3-3 - P&WA Fuel Specification,
Distillate Fuel, Marine and Industrial Gas Turbine Engine

Property	ASTM Test Method	Limits
Distillation Temp, °F	D-86	
IBP		To be reported
10% Evaporation		440 max
20% Evaporation		To be reported
50% Evaporation		675 max
90% Evaporation		725 max
Flash Point, °F	D-93	110 min or legal
Pour Point, °F	D-97	To be reported
Cloud Point, °F	D-97	To be reported
Viscosity, Cs at 100°F	D-445	3.0 max
Carbon Residual (10% Bottoms), wt %	D-524	0.15 max
Sulfur, wt %	D-129	1.0 max
Corrosion at 212°F, ASTM Code No.	D-130	1 max
Ash, wt %	D-482	0.005 max
Gravity, °API	D-287	To be reported
Neutrality	D-1093	Neutral
Net ht of Comb., Btu/lb	D-240 or D-2382	To be reported
Luminometer Number	D-1740	25 min
High Temp Stability	D-1660	
Pressure Change, in Hg		12 max
Preheater Dep Code		2 max
Sediment, mg/gal	D-2276	24 max
Free Water Content, Vol %	-	0.01 max
Trace Metal Contaminants, ppm		
Vanadium		0.1
Sodium		0.1
Potassium		0.1
Calcium		0.1
Lead		0.1
Copper		0.02

Table 3-4 Fuel Oil Pretreatment
FCI and Operating Costs,
Conventional System, 300 gpm Fuel Rate
All Figures in $ Thousand

Item	2-Stage	3-Stage	4-Stage
Fixed Capital Investment	1,680	2,050	2,560
Operating Costs (Annual)			
Direct Costs			
Operating Labor (O.L.)	17.5	17.5	17.5
Utilities			
Electricity	43.8	57.8	75.4
Steam	126.1	126.1	126.1
Process Water	4.7	4.7	4.7
Supplies (30% of O.L.)	5.3	5.3	5.3
Wetting Agent	21.6	26.4	26.4
Maintenance (4% of FCI)	67.1	82.0	102.4
Total Direct Costs	286.1	319.8	357.8
Indirect Costs			
Interest & Amortization, Depreciatiton, Taxes, Insurance, License (18% of FCI)	302.1	368.8	461.0
Total Operating Costs	588.2	688.6	818.8
Average Cost per Barrel Water Washing, cents/bbl	19.6	22.9	27.3
Additional Operating Costs			
Vanadium Inhibitor, cents/bbl/ppm Vanadium[a]	0.9	0.9	0.9

[a] Vanadium content can range from 0.1 ppm to 400 ppm. Property data sheets indicate most coal and oil shale derived fuels will contain less than 1 ppm Vanadium. Petroleum resids from heavy crudes can contain up to 400 ppm.

Table 3-5 - Fuel Properties Data Sheets Supplied to Pretreatment Equipment Manufacturing

Listed are table numbers located in the Literature Survey Appendix.

Section 2	Appendix B	(cont'd)
2-1	B-1	B-33
2-2	B-2	B-34
2-3	B-3	B-35
2-4	B-4	B-36
2-5	B-5	B-37
2-6	B-8	B-38
2-6a	B-9	B-39
2-7	B-10	B-40
2-8	B-11	B-41
2-9	B-12	B-42
2-10	B-13	B-43
2-11	B-14	B-44
2-12	B-15	B-45
2-13	B-16	B-46
2-14	B-17	
	B-18	
Section 3	B-19	
3-1	B-22	
3-2	B-23	
3-3	B-24	
3-4	B-30	
3-5	B-31	
3-6	B-32	

Table 3-6 Fuel Oil Pretreatment
Estimated FCI and Operating Costs,
Alternate Centrifugal Contactor System, 300 gpm Fuel Rate
(All Figures in $ Thousand)

Item	1 Contactor System	2 Contactor System
Fixed Capital Investment	1,200	1,725
Operating Costs (Annual)		
Direct Costs		
Operating Labor (O.L.)	17.5	17.5
Utilities		
Electricity	15.1	30.2
Steam	126.1	126.1
Process Water	4.7	4.7
Supplies (30% of O.L.)	5.3	5.3
Maintenance (4% of FCI)	47.9	69.0
Total Direct Costs	216.6	252.8
Indirect Costs		
Interest & Amortization. Depreciation, Taxes, Insurance, License (18% of FCI)	215.5	310.3
Total Operating Costs	432.1	563.1
Average Cost per Barrel Water Washing, cents/bbl	14.4	18.7
Additional Operating Costs		
Vanadium Inhibitor, cents/bbl/ppm Vanadium[a]	0.9	0.9

[a] Vanadium content can range from 0.1 ppm to 400 ppm. Property data sheets indicate most coal and oil shale derived fuels will contain less than 1 ppm Vanadium. A few petroleum resids can contain up to 400 ppm.

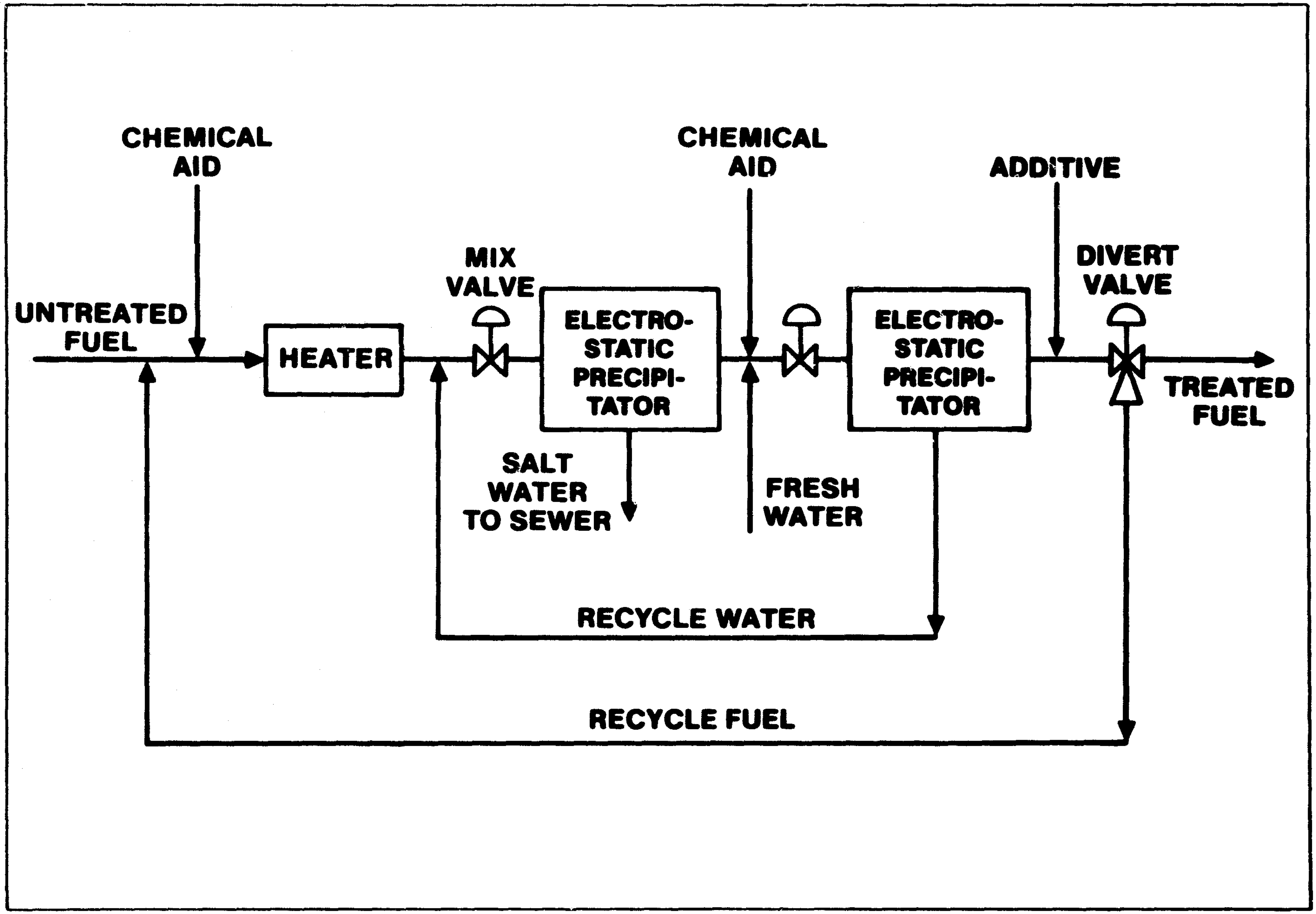

Figure 3-1 - On-Site Pretreatment Block Flow Diagram, Conventional System, Salt Content = 200 ppm

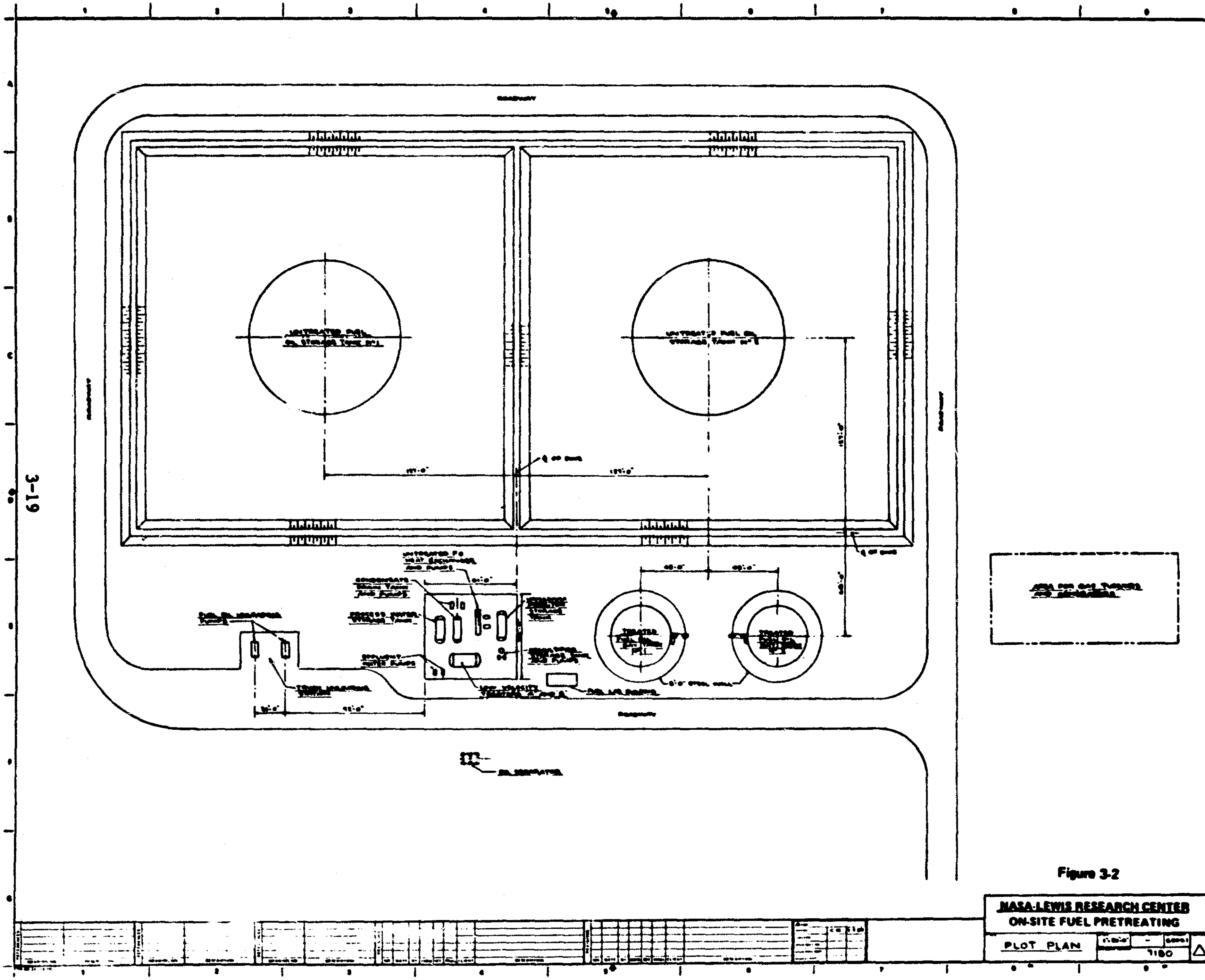
NASA-LEWIS RESEARCH CENTER
ON-SITE FUEL PRETREATING
PLOT PLAN
7180

Figure 3-2

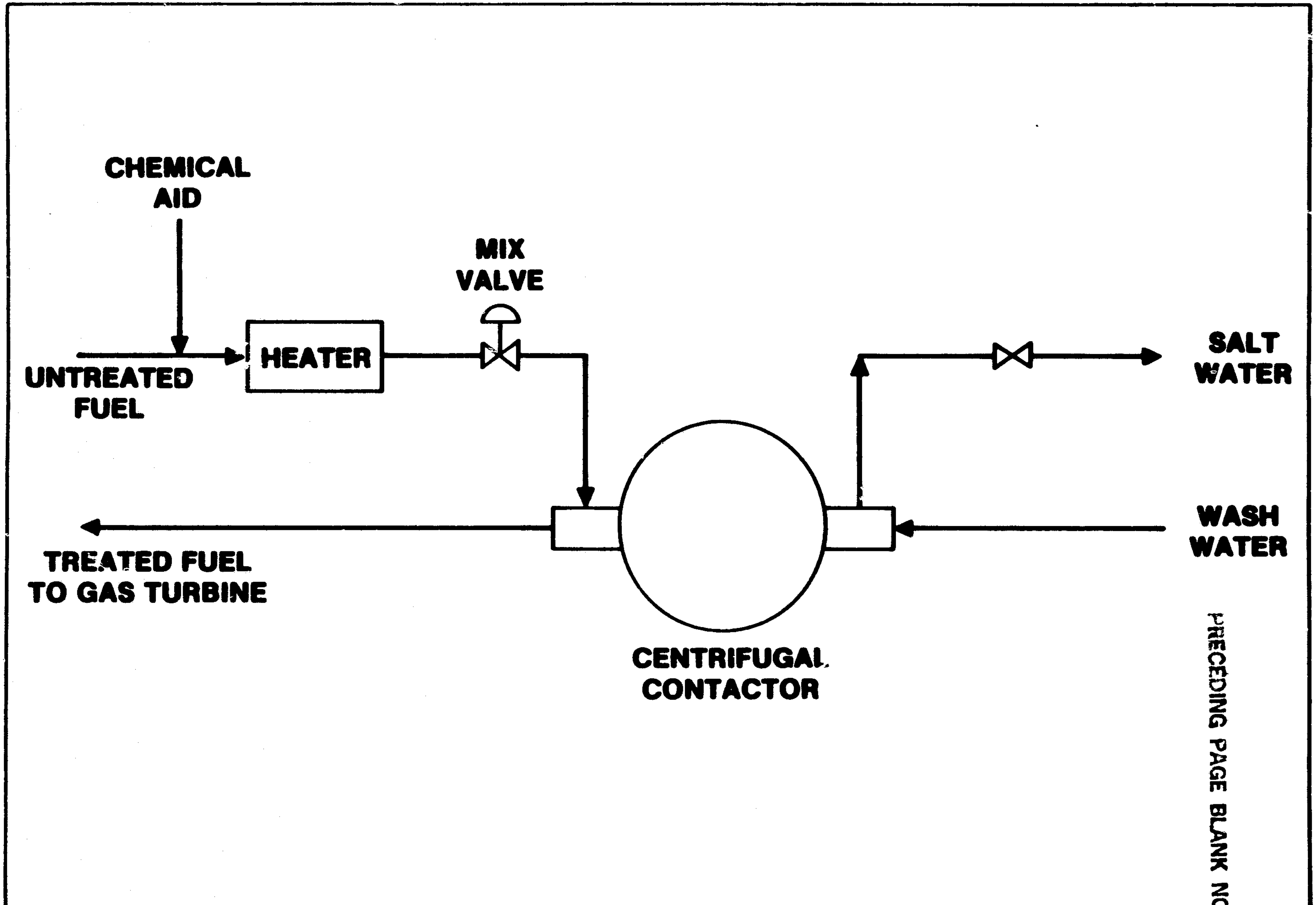

Figure 3-3 - On-Site Pretreatment Block Flow Diagram, Alternate System, Salt Content = 20 - 200 ppm

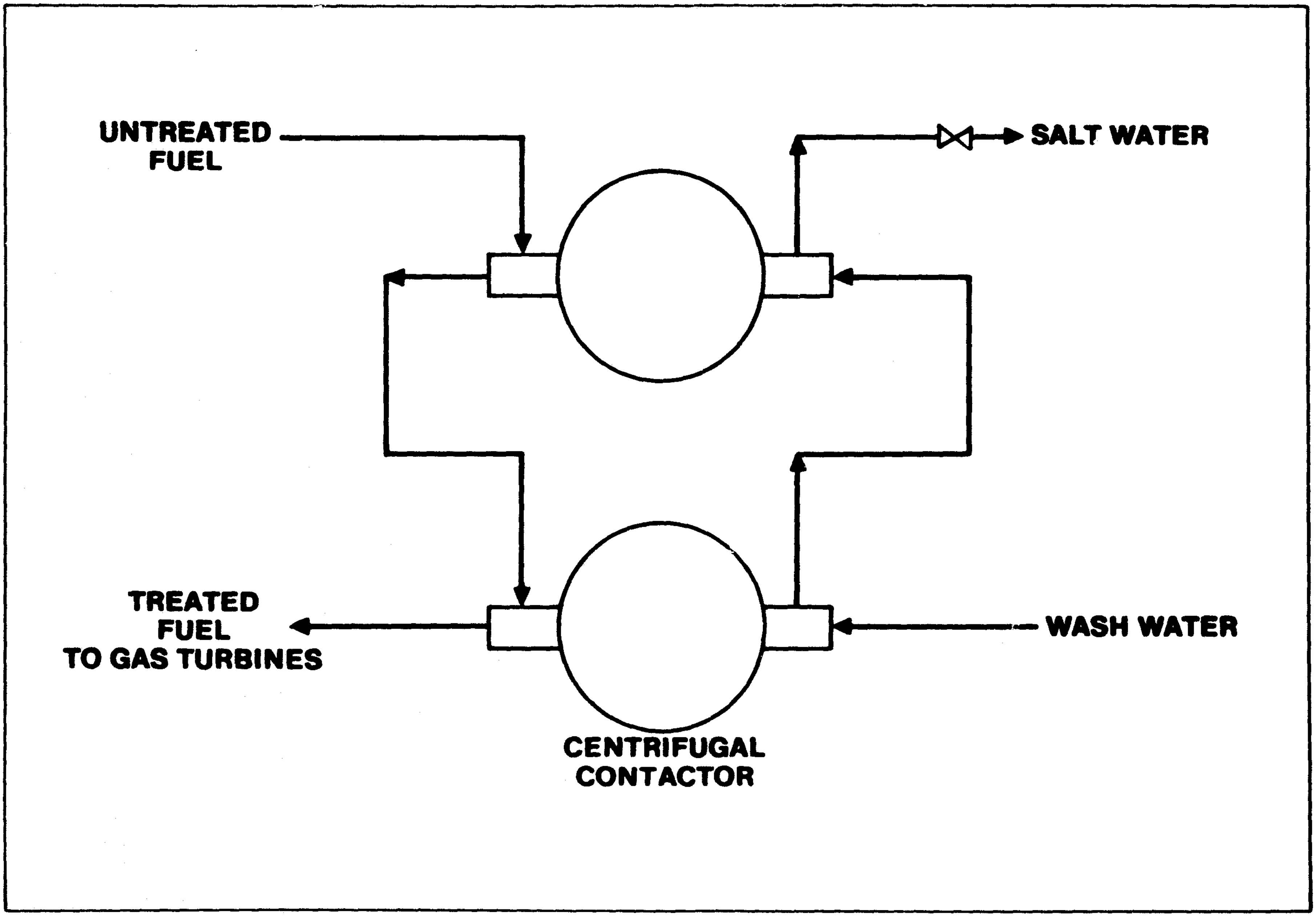

Figure 3-4 - On-Site Pretreatment Block Flow Diagram, Alternate System, Salt Content = 2,000 ppm

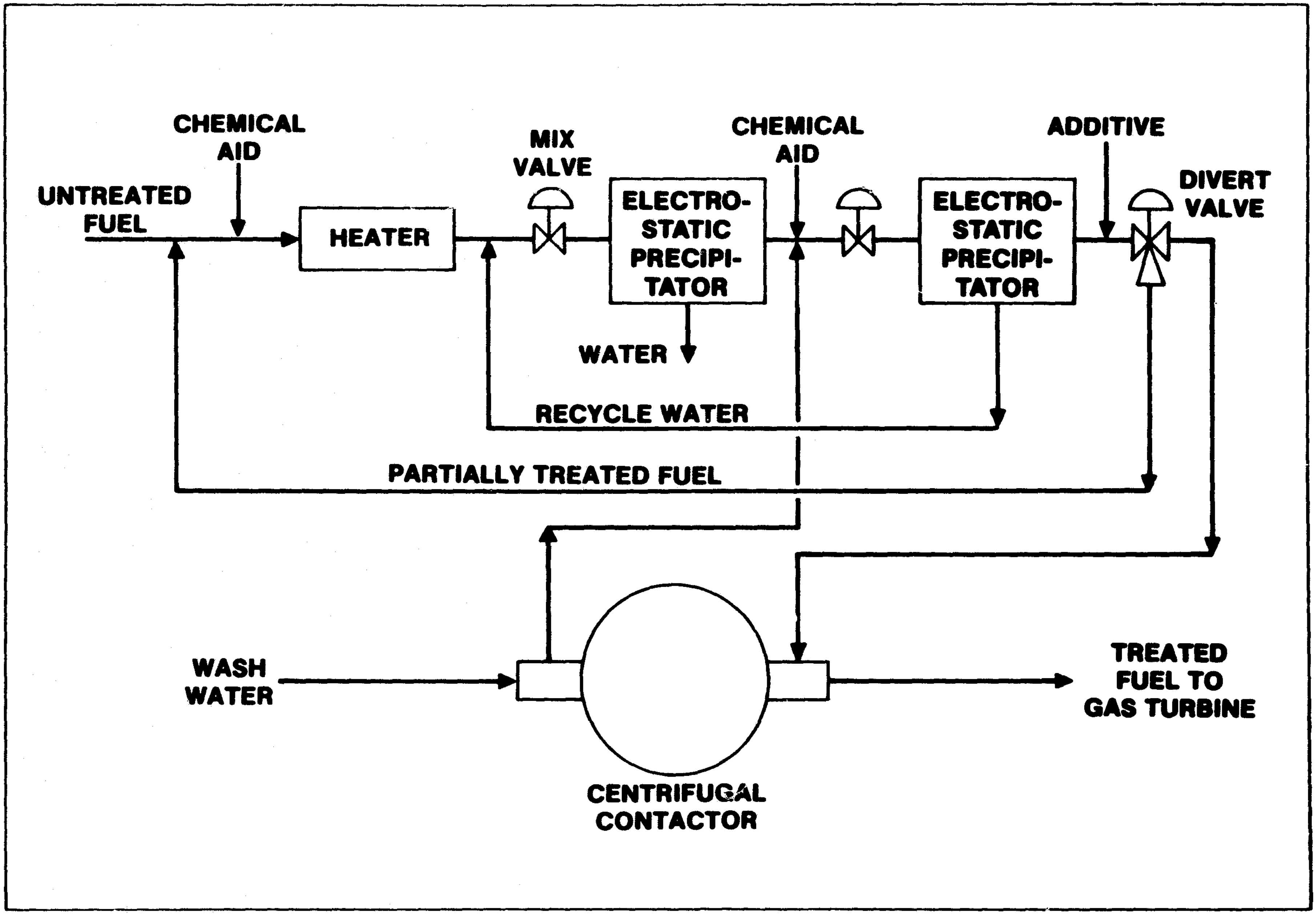

Figure 3-5 - On-Site Pretreatment Block Flow Diagram,
Conventional Plus Alternate System,
Salt Content = 2,000 ppm

SECTION 4

EXISTING REFINERIES TO UPGRADE FUELS

This section presents the results of an investigation of the feasibility and relative costs of upgrading oil shale derived syncrudes and coal derived syncrudes in an existing refinery complex. To achieve the task objectives, a typical U.S. Midwest refinery having a capacity of 200,000 BPD, processing a 60/40 volume percent mixture of South Texas/Light Arabian crudes was selected.

Linear program (LP) model development, cost and process data generation and results obtained are described. A copy of representative results from an individual computer run is included as Exhibit 4-A at the end of this section.

4.1 REFINERY MODEL

The objective in use of a refinery model is to allow a linear program to select the optimum path to produce a given product slate. The optimized refinery output becomes a base case refinery for determining the relative costs of upgrading coal syncrudes and shale syncrudes in an existing refinery by adding necessary expansion units.

The scope of work involves crude selection, product slate selection, refinery unit selection, calculation of process unit yields, determination of physical property data, development of investment and operating costs, definition of product specifications, and establishment of program files for the linear program.

4.1.1 SELECTION OF REFINERY MODEL

The selection of a refinery model was based on an analysis of in-house refinery projects and a literature review. The final selection was based on the Annual Refining Survey, which appeared in "Oil and Gas Journal", March 26, 1979.

Two refineries identified as typical modern day complexes were the Marathon Oil Company refinery at Robinson, Illinois, and the Mobil Oil Corporation refinery at Joliet, Illinois. Both refineries utilize about 200,000 barrels per stream day (BPSD) crude capacity.

The following process units were included in the refinery model based on an analysis of the Marathon and Mobil refinery configurations:

- Crude Unit
- Vacuum Unit
- Naphtha Hydrotreating Units
- Distillate Hydrotreating Unit
- Gas Oil Hydrotreating Unit
- Fluid Catalytic Cracking Unit (FCC)
- Distillate Hydrocracking Unit
- Catalytic Reforming Unit
- Alkylation Unit
- Delayed Coking Unit
- Sulfur Recovery Plant
- Waste Water Treating Plant

4.1.2 INPUT TO REFINERY MODEL

A. Crude Oil Feed

Crude oil feed to the refinery LP model is shown in Table 4-1 as a 60/40 vol % mix of South Texas and Light Arabian crudes. It is based on importing foreign crude oil in a quantity adequate to meet the product volume consumed in the U.S.A., that is, about 40% imported foreign crude. The South Texas domestic crude and the Light Arabian crude analyses were taken from inhouse data sources.

B. Product Slate, Specifications and Product Values

Product slate and specifications applied to the refinery LP model are shown in Table 4-2 along with designated product values in $/bbl.

Product values represent selling price at the refinery gate. The product slate for the existing refinery is also shown on Figure 4-1.

Table 4-2 also includes the refinery feed purchase prices. The shale oil price of $25 per barrel is a rounded average of published costs for shale oil from surface processed underground mined oil shale. The H-Coal syncrude price was placed at $2 per barrel above the $30 petroleum crude price based on a published evaluation wherein H-Coal liquid was estimated by UOP to have a value $2 per barrel greater than a 65/35 Light/Heavy Arabian crude oil blend. The SRC-II syncrude price of $30 per barrel is in the price range cited by the process developer.

The refinery simplified product slate represents the output of a typical refinery. It consists of LPG, non leaded gasoline, distillate fuels and heavy residual fuels, with coke and sulfur as byproducts. The distillate fuels produced are in the specification range of No. 2 fuel oil and the heavy resid meets Midwest market specifications for No. 6 fuel oil. The following major product distribution is chosen to determine the operation of an existing refinery:

LPG	4 Vol % of Crude
Gasoline	54 Vol % of Crude
No. 2 Fuel Oil	27 Vol % of Crude
No. 6 Fuel Oil	10 Vol % of Crude

(Total product volume shown above is not equivalent to total crude volume due to density differences and noninclusion of solid products and fuel gases formed in the processing.)

The gasoline specification is set to meet a 91 research octane number (RON) for non-leaded gasoline with a Reid vapor pressure of 10 psi, maximum.

The specifications and product values were selected to conform to those existing in March, 1980, which is used as the base time period for this analysis. Product values were taken from the Oil and Gas Journal as averages for the first six months of 1980 to conform with the March, 1980 base period.

C. Utility Data

The total refinery energy requirements, i.e. fuel, electricity, etc., are provided by refinery products such that the refinery operation is autonomous except for make-up water. This assures that the utility costs and crude and product costs are consistent. The utility requirements are based on providing a 1250 psig steam plant for driving let down turbines to provide power requirement and low level process steam. Fuel for firing heaters and boiler facilities is supplied from refinery fuel gas and fuel oil generated internally. Cooling water circulation, condensate recovery, and sour water stripping facilities are also provided.

D. Investment Cost Data

Investment cost data for refinery process units are based on in-house estimates prepared by the cost estimating group for a 200,000 BPSD refinery processing South Texas crude. The reference date for all cost data is March, 1980. Capacity ratio exponents used were based on past experience with similar refinery process unit costs.

Royalty, catalyst and chemical requirements cost data are based on in-house data for similar process refinery units.

E. Operating Cost Data

Operating cost data is based on chemical, catalyst, and water usage. Chemical usage is from "Guide to Refinery Costs" W.L. Nelson, 1976. Chemical costs were taken from "Chemical Marketing Reporter" publication. Catalyst costs and usage are based on data for process refinery units.

F. Product Slate

The refinery product slate is based on maximum gasoline production while providing fuel oiils, No. 2 and No. 6, for use in home heating and as boiler fuel in utility plants. Coke and LPG are also produced.

No. 6 fuel oil is produced in limited amounts by blending hydrotreated vacuum gas oil with lighter products since a resid hydrodesulfurization unit is not provided.

4.1.3 OUTPUT OF REFINERY LP MODEL

A. Refinery Optimized Path

The optimized refinery path is shown in Figure 4-1 and represents the existing refinery configuration to be used as the basis for syncrude upgrading. Several aspects of the existing refinery are important to the development of the syncrude upgrading and are listed as follows:

1. The refinery configuration shown in Figure 4-1 sets the process units size, which will remain fixed, in the syncrude upgrading.

2. The refinery requires no hydrogen plant since adequate hydrogen is available from reforming to meet all hydrotreating and hydrocracking requirements.

3. All fuel requirements for the refinery are satisfied from fuel gas and oil generated by internal refinery units.

4. All steam and energy requirements are generated in the refinery from available fuels.

5. The product slate shown in Figure 4-1 represents the petroleum based products from the existing model refinery. As syncrude feed is added to the existing refinery, and turbine fuels are produced with varying specification, the No. 2 and No. 6 fuel oils quantities will vary while the gasoline production remains about the same.

6. As syncrude is added to the existing refinery, crude petroleum feed will be reduced to meet a required product slate by utilizing all process units in the most economical manner.

7. When synfuels are added to the existing petroleum refinery feeds and equipment added to the refinery for processing the new feed material the following criteria are observed:

   (a) Gasoline market shall remain unchanged; thus no increase in refinery gasoline production is allowed.

   (b) Where turbine fuels are produced (production fixed at 20,000 BPD) it is considered that these fuels replace in part other fuels dedicated to generation of electrical energy. Thus the existing refinery plus syncrude is required to produce only 8,000 BPD of No. 6 fuel oil compared to the former 20,000 BPD in the petroleum fed refinery. Maximum No. 2 fuel production is set equal to that made by the basic refinery.

   (c) Where product limits cannot be exceeded, the petroleum charge rate is reduced to bring fuel and products into balance. This is considered consistent with the purpose of manufacture of synthetic fuels, namely, to reduce crude oil imports. In this model the crude reduction is in the same ratio of domestic to foreign as the stated base.

   (d) Extensive hydrotreating of the syncrudes and their fractions will be performed. This field, relating to syncrudes, is in a developing stage. The best available published information coupled with judgment based on in-house experience is utilized.

B. Refinery Capital Cost

The existing refinery battery limits process units' fixed capital investment (FCI), Table 4-3, is approximately $400 million based on March 31, 1980 dollars. This represents 65% of the total refinery FCI. Offsites constitute 35% of the total FCI of approximately $610 million.

C. Refinery Profitability

Refinery Profitability is approximately a 15% discounted cash flow (DCF) rate of return. Table 4-3 contains the calculation of the petroleum refinery's operating margin of $702,000 per stream day which consists of the recovery of capital associated costs and a profit approximating a 15% DCF. This operating margin will appear in tables involving co-processing of syncrudes in the existing refinery in the calculation of case required revenues. The economic parameters used in developing the costs include:

- 20-year operation
- Fixed Capital Investment
- Profit - 15% DCF
- Income Tax, 50% using double declining balance depreciation with 16 year useful life
- 10% Investment Tax credit
- 4% of FCI annual maintenance labor and materials
- 2.5% of FCI annual property taxes and insurance costs
- Allowance for spare parts inventory and working capital.

The above capital cost factors amount to 35% of the fixed capital investment. This is applied to the FCI additions and new refinery FCIs for determination of turbine fuel required revenue in each of the cases in the summary tables contained in Section 6.

D. Utility Output

Utility and fuel requirements for the existing refinery operating on petroleum feed are summarized in Table 4-4.

Table 4-1 - Petroleum Crude Feedstock to Existing Refinery

Crude Type: 60/40 Mixed South Texas and Light Arabian

°API	36.9
Sp. Gr.	0.8403
Sulfur, wt %	0.948
Nitrogen, wt %	0.20
Oxygen, wt %	0.03
Metals (Iron, Vanadium, Nickel) ppm, wt	30.0

TBP Analysis:

wt %	°F
ST/10	IBP/210
10/30	210/405
30/50	405/570
50/70	570/775
70/85	775/990
85/EP	990/1,000+ (resid)

Table 4-2 - Refinery Model - Product Characteristics, Feed and Product Values

Non-Leaded Gasoline

Research Octane Number	91 (min)
Reid Vapor Pressure	10 PSI (max)
Product Value	40 $/Bbl

No. 2 Fuel Oil

Viscosity	2 cst @ 100°F (min)
	3.5 cst @ 100°F (max)
°API	30.0 (min)
Sulfur	0.2 wt % (max)
Product Value	32 $/Bbl

No. 6 Fuel Oil

Viscosity	50 cst @ 122°F (min)
	500 cst @ 122°F (max)
°API	Report
Sulfur	1.0 wt % (max)
Product Value	23 $/Bbl

Refinery Fuel Oil

Same as No. 6 Fuel Oil, Except Viscosity 9.5 cst @ 122°F (min)

Feed Values

Petroleum Crude	$30/Bbl
Shale Oil	$25/Bbl
H-Coal	$32/Bbl
SRC-II	$30/Bbl

Turbine Fuel

Viscosity	1.8 cst @ 100°F (min)
	200 cst @ 100°F (max)
°API	15.0 (min)
Sulfur	0.7 wt % (max)
Nitrogen	0.25 wt % (max)

LP calculation at 0 value. Required selling price to be hand computed for each case.

LPG

Propane, butane components
$25/Bbl, Product value

Coke

Heat of combustion (HHV), 30 MMBtu/ton
$20/Ton, Product value

Sulfur

Heat of combustion (HHV) 8.937 MMBtu/Lton
$109.5/Lton, Product value

Ammonia

Heat of combustion (HHV) 19.336 MMBtu/ton
$190/ton, Product Value

Table 4-3 - Refinery Capital Costs and Profitability Analysis

Process Units Fixed Capital Investment:

Capacity (BPD)	Process Units	$ Million
200,000	Crude	44.0
74,900	Vacuum	31.2
60,140	Naphtha HDS	14.9
21,670	Atm Gas Oil HDS	16.2
5,110	Vac Gas Oil HDS	5.6
49,480	FCC	56.0
10,190	Hydrocracker	44.5
12,350	Coker	30.0
48,215	Reformer	49.0
7,930	Alkylation	16.0
132 LTPD	Sulfur Plant	5.2
5,259 M lb/D	SWS Plant	2.2
15,022 M lb/D[a]	Power Plant	75.0
194,000 M Gal/D	CWC Plant	6.4
		396.2

Process units @ 65% of Total FCI, offsites at 35% of Total FCI.

$$\text{Total Fixed Capital Investment} = \frac{396.2 \text{ million}}{0.65} = \$609.5 \text{ million}$$

Daily Operating Margin (Capital Associated Costs and Profit):

Product Value		$6,750,000
Deduct: Feed Cost	$6,000,000	
Operating Cost	47,796	$6,047,796
Operating Margin[b]		$ 702,204

[a] 1250 psig steam, approximately 50 kW electricity generation.

[b] The operating margin is the gross return from operations and consists of the product value less the cost of feed and operating costs. Operating costs in this case include catalysts, chemicals, operating labor and supplies. When expanding the refinery only operating revenues greater than those at the existing refinery are required to provide an equivalent return on the additional investment required for expansion.

Table 4-4 - Refinery Total Utilities Requirement
(Computer Output)

Unit	Usage Rate		
Sour Water Stripping	5,260	M lb/D	(440 gpm)
Cooling Water Circulation	195	MM gal/D	(135,400 gpm)
Power Generation	1,240	M kWh/D	(51,650 kW)
Steam Boiler (1200 psig)	15,020	M lb/D	(626 M lb/hr)
Fuel Consumption	79,098	MM Btu/D (LHV)	(13,183 BPD FOE)

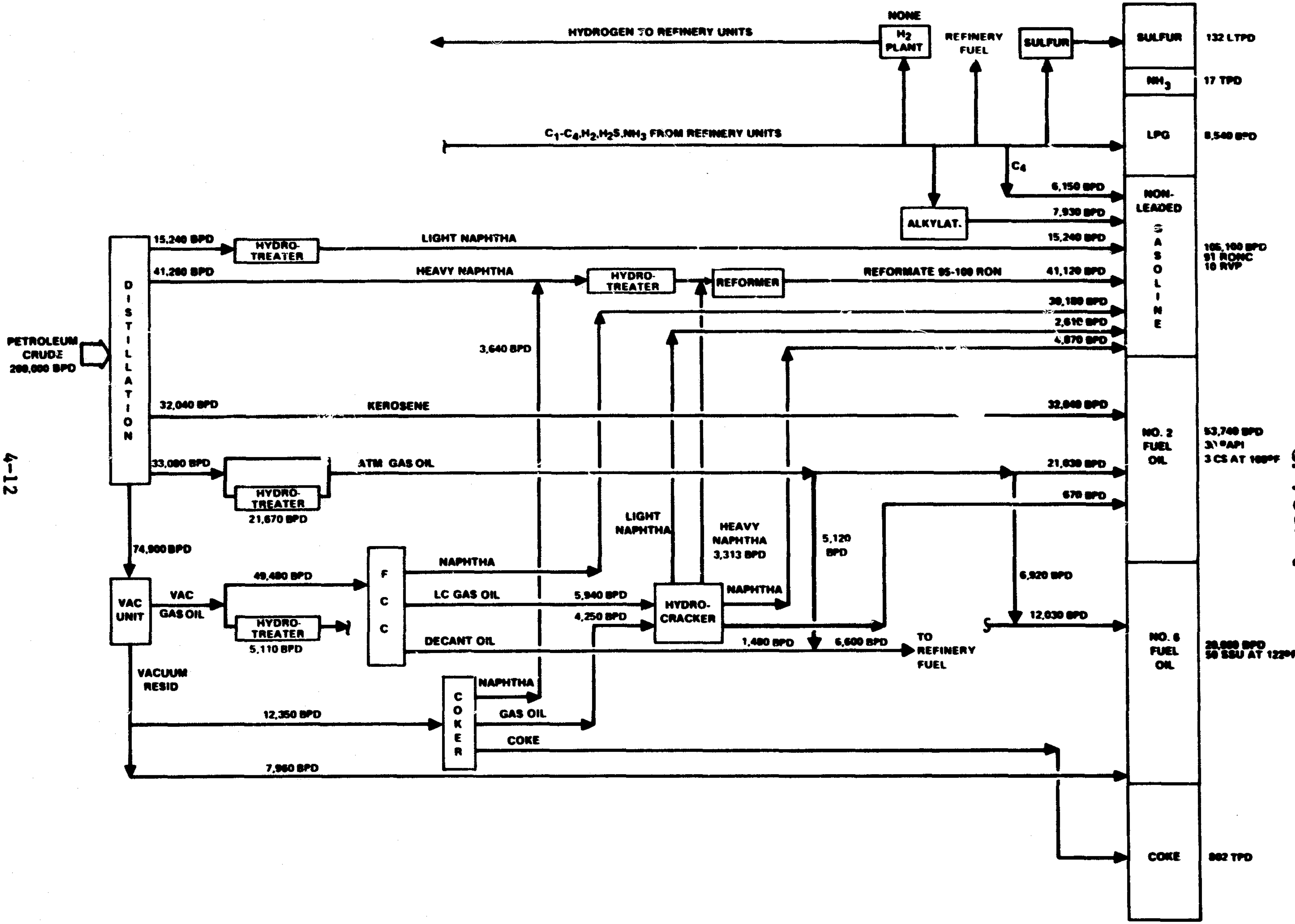

Figure 4-1 - Computer Output Data Diagram, Existing Petroleum Crude Refinery

4.2 SHALE OIL PLUS EXISTING REFINERY

The objective in development of a model of an existing petroleum refinery to process shale oil is to use linear programming to select the optimum economical path to meet a given product slate. The optimized output result will be evaluated for relative costs of upgrading shale oil to turbine fuels and petroleum-grade products.

The scope of work involves shale oil feed selection, product slate selection, shale oil process path configurations, calculation of process unit yields, determining physical property data, obtaining cost and operating values, definition of product specifications, and establishing program files for entry to the LP program.

4.2.1 SHALE OIL MODEL

The selection of a shale oil model is based on pilot plant work carried out and still underway by Chevron Research Company as performed under Contract No. EX-76-C-01-2315 for the U.S. Department of Energy. Two process paths were selected as potential routes for economic evaluation as follows:

a. Hydrotreating the whole shale oil syncrude to a low nitrogen level before distillation with subsequent upgrading in process units to petroleum grade products, and

b. Hydrotreating individual cuts after distillation to a low nitrogen level with subsequent upgrading in process units to petroleum grade products.

4.2.2 INPUT TO SHALE OIL MODEL

A. Shale Oil Input Diagrams

The computer input paths are shown in Figures 4-2 and 4-3 which represent the configurations used as the basis for shale oil upgrading

and economic evaluation for a given product slate. The configurations show process yield data, hydrogen consumption, stream names, process units, and optional paths for optimization.

Figure 4-2, hydrotreating before distillation, depicts two-stage hydrodenitrification of whole shale oil to a nitrogen level of about 550 ppm (wt.). The low pressure first-stage hydrotreating serves to saturate the olefinic molecules resulting from the pyrolysis of shale oil and to remove metals such as arsenic and iron which are potential catalyst poisons. The low pressure stage, or guard bed, may be located either in the production facility, if hydrotreating is required to prevent polymerization in the pipeline, or at the refinery if the transportation problem can be overcome. The availability of a hydrogen source at the refinery enhances the economics of this location.

In the second stage high pressure hydrotreater, hydrodenitrification occurs along with considerable upgrading of the 650° F plus fraction which results in an excellent feed for FCC and hydrocracking processes. The naphthas from the hydrotreating operation are upgraded to high octane gasolines by catalytic reforming. The middle distillate fractions may require additional hydrotreating to produce diesel or jet fuels, or may be bypassed around the hydrotreater to make No. 2 fuel oil. There is essentially no residuum boiling above 1000°F available, the heaviest fraction being the 650-950°F cut, which eliminates the need for resid hydrotreating. The process units that would be required for upgrading shale oil as shown in Figure 4-2, are as follows:

- Low pressure Hydrotreater
- High pressure Hydrotreater
- Distillation
- Naphtha Hydrotreater
- Distillate Hydrotreater
- Catalytic Reformer
- Hydrocracker (single-stage)
- Catalytic Cracker

Figure 4-3, hydrotreating after distillation, is based on single stage stabilization and hydrotreating of whole shale oil with about 30 percent nitrogen removal to a nitrogen level of about 1.4 wt%. The process units involved are as follows:

- o Low pressure hydrotreater
- o Distillation
- o Naphtha Hydrotreater
- o Distillate Hydrotreater
- o Catalytic Reformer
- o Hydrocracker (two-stage)
- o Catalytic Cracker
- o FCC Hydrotreater
- o Heavy Fuel Hydrotreater

The individual fractions from low pressure stabilization must be hydrotreated in high pressure, low space velocity reactors to reduce the nitrogen to the low levels required to prevent poisoning and deactivation of the catalyst in the subsequent processing units listed above. After hydrotreating, these fractions will be upgraded to high octane gasolines, No. 2 fuel oil, No. 6 fuel oil and turbine fuels.

It should be pointed out that the configurations shown in Figures 4-2 and 4-3 are to be combined with the existing refinery, and as such, may be using existing refinery unit capacity or may be adding new capacity to existing refinery units. The LP model will use this option in the combined refinery in selecting the optimum path to meet a given product slate.

It should also be pointed out that hydrotreating whole shale oil or shale oil fractions in Figures 4-2 and 4-3 requires special reactors (high pressure and low space velocities) and that these will be new units in the shale oil refinery.

B. <u>Shale Oil Feed</u>

The shale oil feed characteristics used as input data to the refinery model is dewatered Paraho shale oil produced from an indirectly heated mode as shown in Table 4-5.

C. <u>Product Slate, Specifications and Values</u>

Product slate and characteristics for the shale oil LP model, the same as for petroleum products, are shown in Table 4-2 along with designated product values in $/bbl. Turbine fuel specifications are also shown in Table 4-2.

D. <u>Investment Cost Data</u>

Investment cost data for shale oil hydrotreating process units was based on in-house estimates. These estimates provide for the pressure levels and space velocities used to treat the raw shale oil. The reference data for all cost data is March 31, 1980, capacity ratio exponents (powers) were based on past experience with refinery unit costs.

Royalty and cost data are based on in-house data for process refinery units.

E. <u>Operating Cost Data</u>

Operating cost data is based on chemical, catalyst, water usage and labor. Chemical usage is from "Guide to Refinery Costs," W. C. Nelson, 1976. Chemical costs were taken from "Chemical Marketing Reporter" publication. Catalyst costs and usage are based on data for process refinery units.

Table 4-5 - Raw Shale Oil Feed

Properties

°API	21.4
Sulfur, wt%	0.6
Nitrogen, wt%	2.0
Carbon, wt%	84.8
Hydrogen, wt%	11.4
Oxygen, wt%	1.3
Arsenic, ppm	12.0
Iron, ppm	33.0
Vanadium, ppm	0.2
Nickel, ppm	2.0
Sodium, ppm	1.4
Viscosity @122°F, cst.	17.0
Viscosity @210°F, cst.	7.0
Pour point, °F	+85

Composition

C_5-350°F Naphtha	6.0
350-650°F Distillate	26.0
650°F+ Bottoms	68.0
	100.0

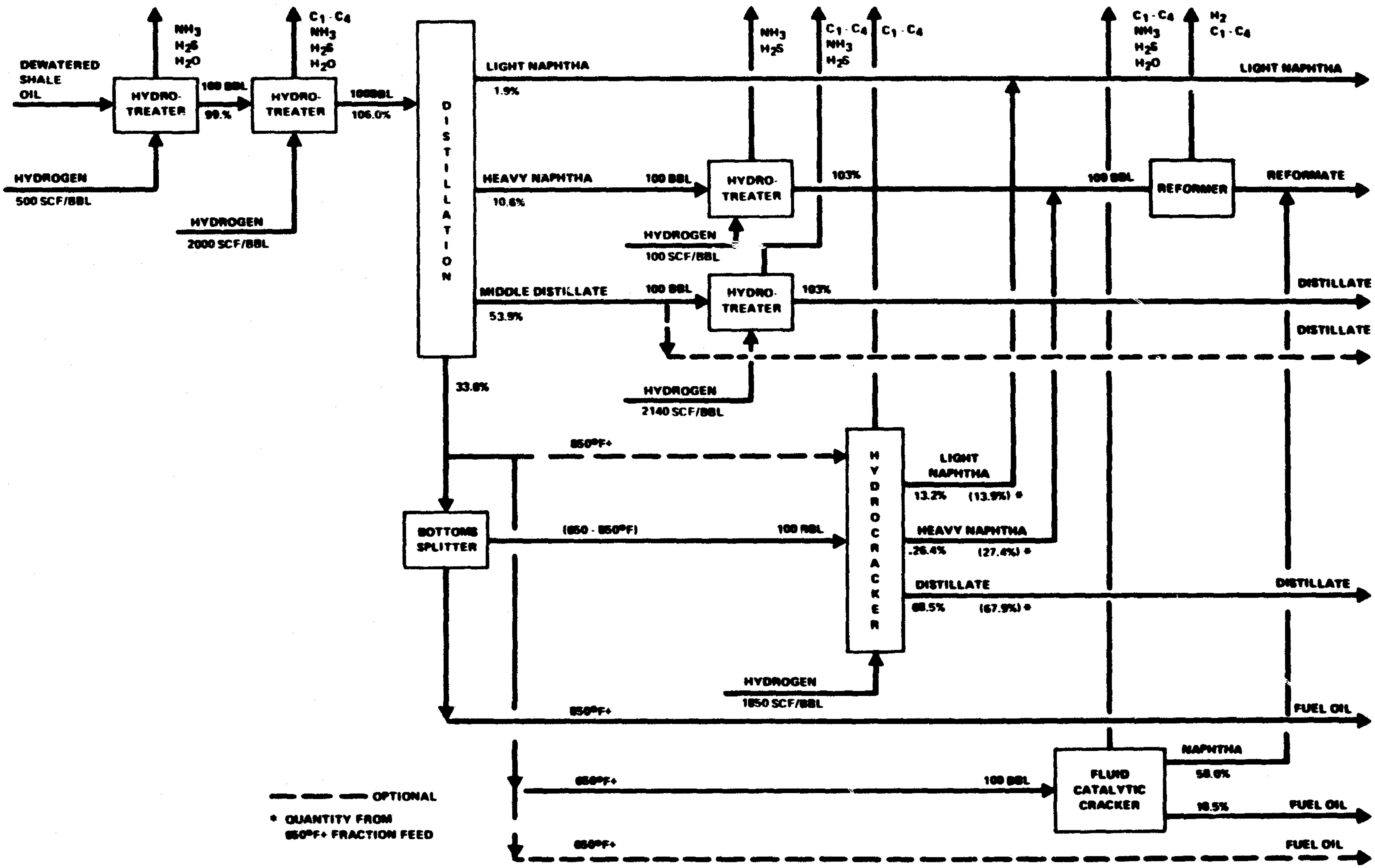

Figure 4-2 - Computer Input Data Diagram, Shale Oil Refining, Case 1

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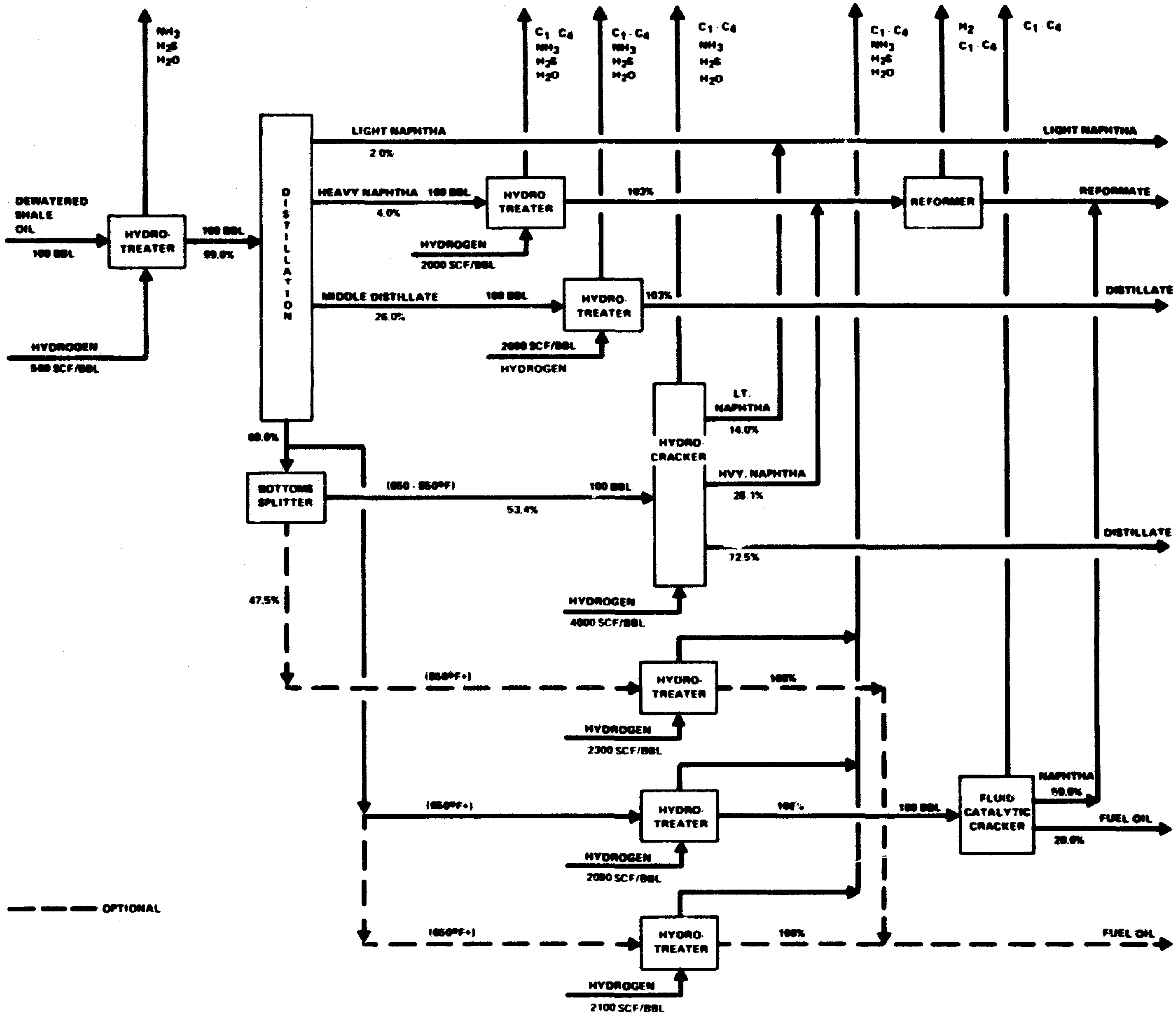

Figure 4-3 - Computer Input Data Diagram, Shale Oil Refining, Case 2

4.3 H-COAL OIL PLUS EXISTING REFINERY

The objective in development of a model of an existing petroleum refinery to process H-Coal syncrude is to allow linear programming to select the optimum economical path to meet a given product slate. The optimized output result will be evaluated for relative costs of upgrading H-Coal to turbine fuels and petroleum-grade products.

The scope of work involves H-Coal oil feed selection, product slate selection, H-Coal oil process path configurations, calculation of process unit yields, determining physical property data, obtaining cost and operating values, definition of product specifications, and establishing program files for entry to the LP program.

4.3.1 H-COAL OIL MODEL

The selection of the H-Coal oil model is based on the following sources:

(1) "Analytical Studies for the H-Coal Process," Mobil Research and Development Corporation, November 28, 1978, performed under Contract No. EF-77-6-01-2676 for the U.S. Department of Energy.

(2) "Crude Oil Versus Coal Oil Processing, Comparison Study," UOP, Incorporated, August 22, 1979, performed under Contract No. EF-77-C-01-2566 for the U.S. Department of energy.

Two process paths were selected as potential routes for economic evaluation as follows:

(1) Hydrotreating the whole H-Coal oil syncrude to a low nitrogen level before distillation with subsequent upgrading in process units to petroleum grade products, and

(2) Hydrotreating individual cuts <u>after distillation</u> to a low nitrogen level with subsequent upgrading in process units to petroleum grade products.

4.3.2 INPUT TO H-COAL MODEL

A. H-Coal Oil Input Diagrams

The computer input paths are shown in Figures 4-4 and 4-5 which represent the configurations used as the basis for H-Coal upgrading and economic evaluation for a given product slate. The configurations show process yield data, hydrogen consumption, stream names, process units, and optional paths for optimization.

In Figure 4-4, hydrotreating before distillation is based on single stage hydrodenitrification of H-Coal oil syncrude to a nitrogen level of about 50 ppm (wt). During the hydrodenitrification reaction, considerable upgrading of the 550°F plus fraction occurs which results in an excellent feed for the fluid catalytic cracking process. The naphthas from the hydrotreating operation are upgraded to high octane gasolines by catalytic reforming. The 350°F plus fraction makes an excellent feedstock for the hydrocracking process. The middle distillate (350-550°F) fraction may be sent directly to No. 2 fuel oil blending, while the 550°F plus fraction may be bypassed around the FCC unit to fuel oil blending.

In the H-Coal cases, there is essentially no residuum boiling above 900°F, the heaviest fraction being the 550-900°F cut, which eliminates the need for resid hydrotreating. All heavier fractions produced in the H-Coal process are either recycled or used for hydrogen production in the liquefaction process. The process units that would be required for upgrading H-Coal, as shown in Figure 4-4, are as follows:

- High pressure hydrotreater
- Distillation
- Naphtha Hydrotreater

- Distillate Hydrotreater
- Catalytic Reformer
- Hydrocracker (single-stage)
- Catalytic Cracker

In Figure 4-5, hydrotreating after distillation is based on distillation of the H-Coal syncrude followed by hydrotreating of the individual fractions. The process units involved are as follows:

- Distillation
- Naphtha Hydrotreater
- Distillate Hydrotreater
- Catalytic Reformer
- Hydrocracker (two-stage)
- Catalytic Cracker
- FCC Hydrotreater
- Heavy Fuel Hydrotreater

The individual fractions from distillation must be hydrotreated in high pressure, low space velocity reactors to reduce the nitrogen to the low levels required to prevent poisoning and deactivation of the catalyst in the subsequent processing units listed above. After hydrotreating, these fractions will be upgraded to high octane gasolines, No. 2 fuel oil, No. 6 fuel oil and turbine fuels.

It should be pointed out that the configurations shown in Figures 4-4 and 4-5 are to be combined with the existing refinery, and as such, may be using existing refinery unit capacity or may be adding new capacity to existing refinery units. The LP model will use this option in the combined refinery in selecting the optimum path to meet a given product slate.

It should also be pointed out that hydrotreating whole H-Coal syncrude or H-Coal oil fractions in Figures 4-4 and 4-5 requires special

reactors (high pressure and low space velocities) and that these will be new units in the H-Coal oil refinery.

B. H-Coal Oil Feed

The H-Coal oil feed used as input to the computer from the liquefaction process is shown in Table 4-6.

C. Product Slate, Specifications and Values

Product slate and specifications (characteristics) for the H-Coal oil LP model, the same as for petroleum products, are shown in Table 4-2 along with designated product values in $/bbl. Turbine fuel characteristics are the same as shown in Table 4-2.

D. Investment Cost Data

Investment cost data for H-Coal oil hydrotreating process units is based on in-house estimates. These estimates are factored to the pressure levels and space velocities used to treat the H-Coal oil. The reference data for all cost data is March 31, 1980. Capacity ratio exponents (powers) were based on past experience with refinery unit costs.

Royalty and cost data are based on in-house data for process refinery units.

E. Operating Cost Data

Operating cost data is based on chemical, catalyst, water usage and labor. Chemical usage is from "Guide to Refinery Costs," W. C. Nelson, 1976. Chemical costs were taken from "Chemical Marketing Reporter" publication. Catalyst costs and usage are based on data for process refinery units.

Table 4-6 - H-Coal Oil Feed

Properties

°API	30.5
Sulfur, wt%	0.15
Nitrogen, wt%	0.37
Carbon, wt%	86.7
Hydrogen, wt%	11.0
Oxygen, wt%	1.72
Nickel, ppm-wt	1.0
Vanadium, ppm-wt	1.0
Arsenic, ppm-wt	0.5
Viscosity @ 122°F, cst	4.2
Viscosity @ 210°F, cst	1.7
Pour point, °F	-45

Composition

		Volume %
IBP-350°F	Naphtha	45.0
350-550°F	Distillate	42.0
550°F+	Bottoms	13.0
		100.0

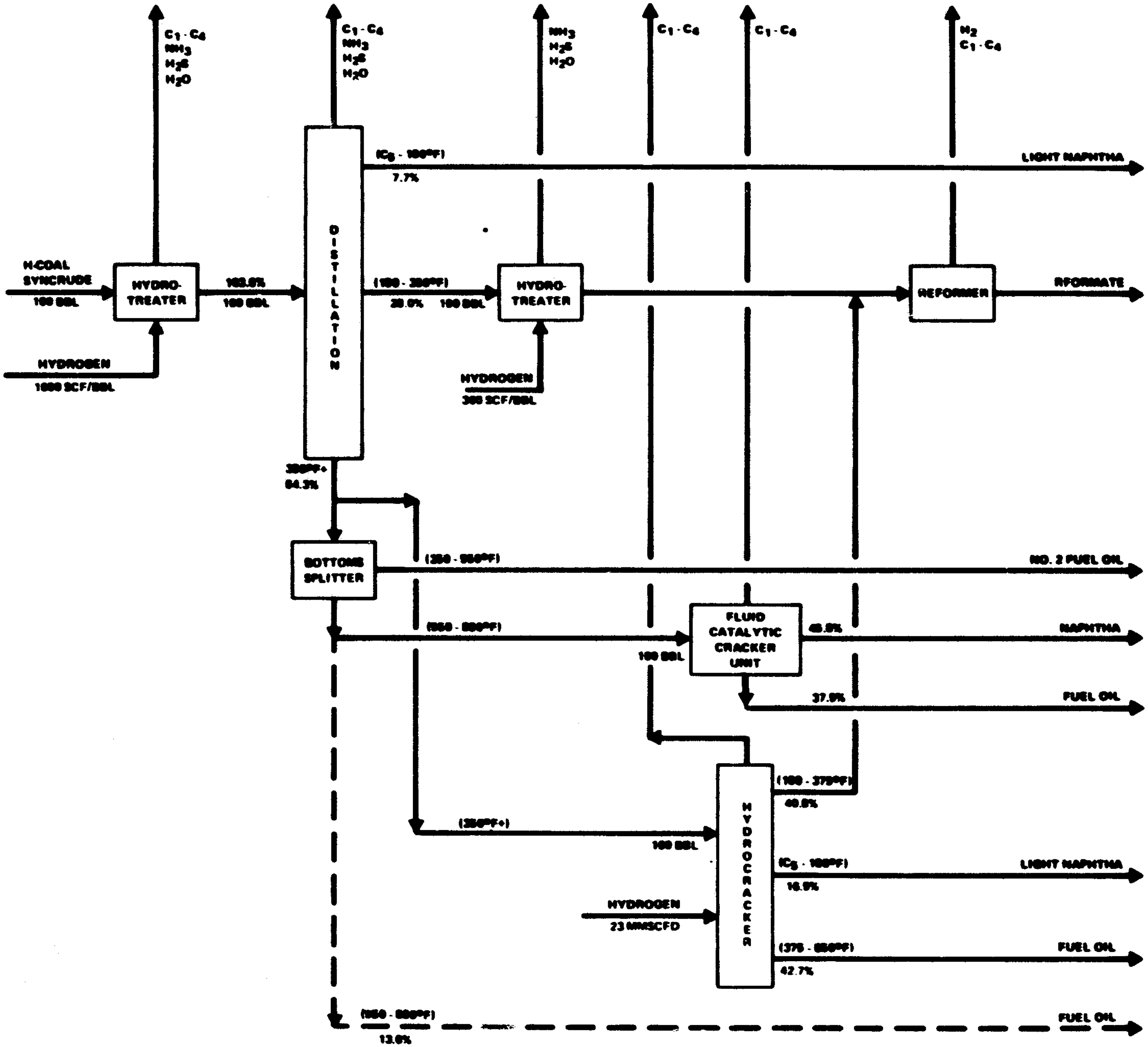

Figure 4-4 - Computer Input Data Diagram, H-Coal Refining, Case 1

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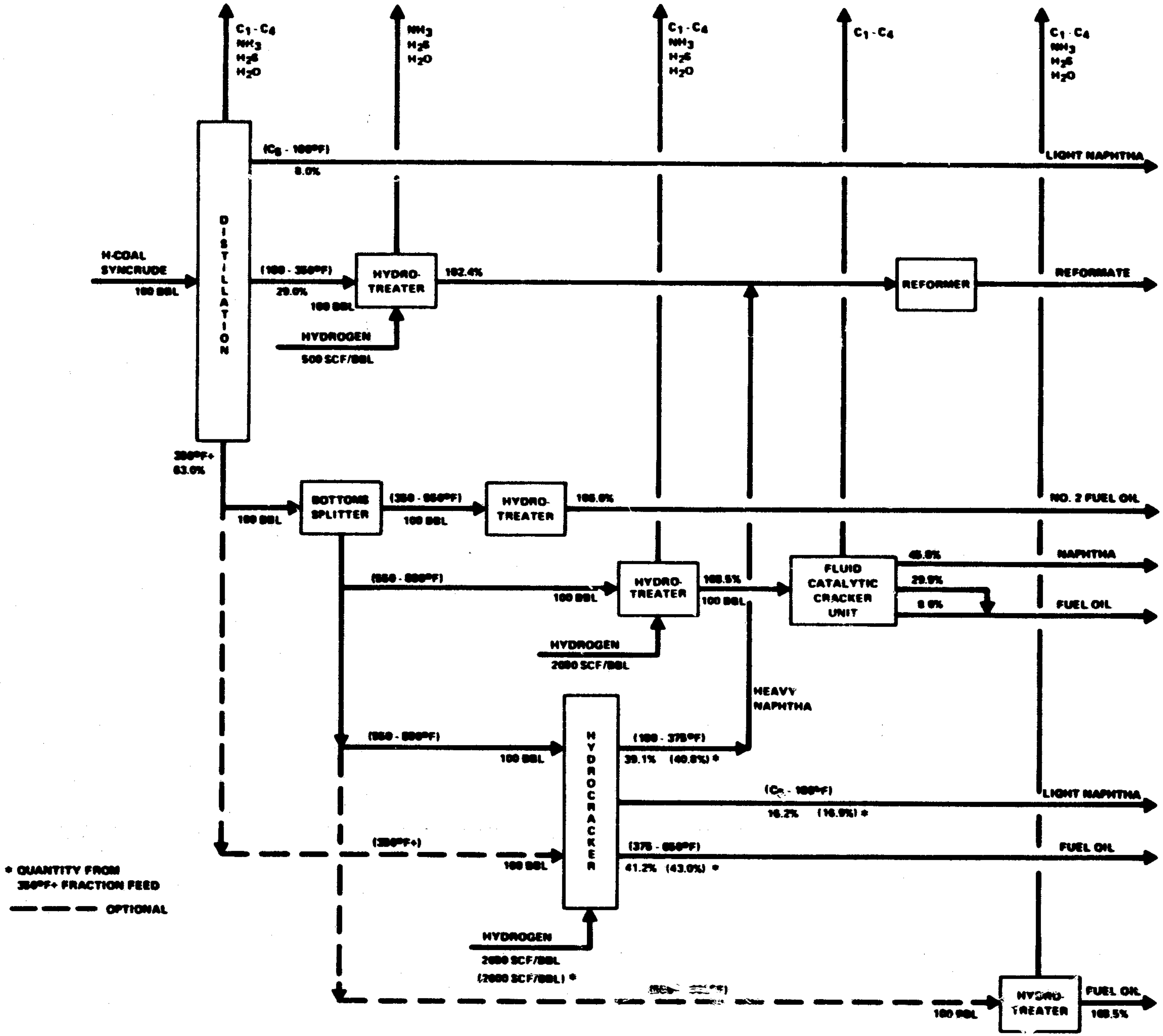

Figure 4-5 - Computer Input Data Diagram, H-Coal Refining, Case 2

4.4 SRC-II OIL PLUS EXISTING REFINERY

The objective in an SRC-II oil plus existing petroleum refinery model is to use linear programming to select the optimum economical path to meet a given product slate. The optimized output results will be evaluated for relative costs of upgrading SRC-II oil to turbine fuels and petroleum-grade products.

The scope of work involves SRC-II oil feed selection, product slate selection, SRC-II oil process path configurations, calculation of process unit yields, determination of physical property data, obtaining cost and operating values, setting of product specifications, and establishing program files for entry to the LP program.

4.4.1 SRC-II OIL MODEL

The selection of an SRC-II oil model is based on pilot plant work carried out by Chevron Research Company under Contract EF-76-C-01-2315 for the U.S. Department of Energy as published in several different quarterly reports during the period 1978 and 1979, as follows: "Refining and Upgrading of Synfuels from Coal and Oil Shales By Advanced Catalytic Processes," R. F. Sullivan *et al*., FE-2315-31, 34, 37, and 40, Chevron Research Company, Richmond, California.

Two process paths were selected as potential routes for economic evaluation as follows:

(1) Separation of the 950°F plus fraction by distillation, hydrotreating the SRC-II oil 950°F minus fraction to a low nitrogen level *before distillation* with subsequent upgrading in process units to petroleum grade products, and

(2) Separation of the 950°F plus fraction by distillation, hydrotreating individual cuts of 950°F minus fraction *after distillation* to a low nitrogen level with subsequent

upgrading in process units to petroleum grade products. The 950°F plus fraction is also upgraded to petroleum grade products.

4. .2 INPUT TO SRC-II OIL MODEL

A. SRC-II Oil Input Diagrams

The computer input paths are shown in Figures 4-6 and 4-7 which represent the configurations used as the basis for SRC-II oil upgrading and economic evaluation for a given product slate. The configurations show process yield data, hydrogen consumption, stream names, process units, and optional paths for optimization.

In the SRC-II cases, the syncrude from the liquefaction process contains the 950°F plus residuum fraction. Unlike the H-coal process where the heavy bottoms fraction was recycled or used for hydrogen production, in the SRC process the bottoms fraction is upgraded to petroleum products.

In the feed to the vacuum distillation, Figure 4-6, there is about 49 volume % residuum boiling above 950°F which requires further processing by delayed coking with coker product hydrotreating to petroleum grade products. Some 950°F plus resid fraction may bypass the coker to fuel oil blending.

In Figure 4-6, hydrotreating before distillation is based on single stage hydrodenitrification of the SRC-II 950°F and lighter fraction from vacuum distillation to a nitrogen level of about 350 ppm (wt). During the hydrodenitrification reaction, considerable upgrading of the 550°F plus fraction occurs which results in an excellent feed for the fluid catalytic cracking process. The naphthas from the hydrotreating operation are upgraded to high octane gasolines by catalytic reforming. The 350°F plus fraction makes an excellent feedstock for the hydrocracking process. The middle distillate (350-550°F) fraction may be sent directly to No. 2 fuel oil

blending, while the 550°F plus fraction may be bypassed around the FCC unit to fuel oil blending. The process units that would be required for upgrading SRC-II in Figure 4-6 are as follows:

- Distillation (vacuum and atmospheric)
- High pressure hydrotreater
- Naphtha hydrotreater
- Distillate hydrotreater
- Catalytic reformer
- Hydrocracker (single-stage)
- Catalytic cracker
- Delayed coker
- Coker product hydrotreaters

In Figure 4-7, hydrotreating after distillation is based on distillation of the SRC-II 950°F- fraction followed by hydrotreating of the individual fractions. The process units involved are as follows:

- Distillation (vacuum and atmospheric)
- Naphtha hydrotreater
- Distillate hydrotreater
- Catalytic reformer
- Hydrocracker (single-stage)
- Delayed coker
- Coker product hydrotreaters
- 400°F+ hydrotreater

The individual fractions from distillation must be hydrotreated in high pressure, low space velocity reactors to reduce the nitrogen to the low levels required to prevent poisoning and deactivation of the catalyst in the subsequent processing units listed above. After hydrotreating, these fractions will be upgraded to high octane gasolines, No. 2 fuel oil, No. 6 fuel oil and turbine fuels.

It should be pointed out that the configurations shown in Figures 4-6 and 4-7 are to be combined with the existing petroleum crude refinery, and as such, may be using existing refinery unit capacity or may be adding new capacity to existing refinery units. The LP model will use this option in the combined refinery in selecting the optimum path to meet a given product slate.

It should also be pointed out that hydrotreating SRC-II oil fractions in Figures 4-6 and 4-7 requires special reactors (high pressure and low space velocities) and that these will be new units in the SRC-II oil refinery.

B. SRC-II Oil Feed

The SRC-II oil feed used as input to the computer from the liquefaction process is shown in Table 4-7.

C. Product Slate, Specifications and Values

Product slate and specifications (characteristics) for the SRC-II oil LP model, the same as for petroleum products, are shown in Table 4-2 along with designated product values in \$/bbl. Turbine fuel characteristics are the same as shown in Table 4-2.

D. Investment Cost Data

Investment cost data for SRC-II oil hydrotreating process units is based on in-house estimates. These estimates provide for the pressure levels and space velocities used to treat the SRC-II oil. The reference date for all cost data is March, 1980. Capacity ratio exponents (powers) were based on past experience with refinery unit costs.

Royalty and cost data are based on in-house data for process refinery units.

E. Operating Cost Data

Operating cost data is based on chemical, catalyst, water usage and labor. Chemical usage is from "Guide to Refinery Costs," W. C. Nelson, 1976. Chemical costs were taken from "Chemical Marketing Reporter" publication. Catalyst costs and usage are based on data for process refinery units.

Table 4-7 - SRC-II Oil Feed

Properties

°API	5.2
Sulfur, Wt %	0.4
Nitrogen, Wt %	1.2
Carbon, Wt %	84.6
Hydrogen, Wt %	9.1
Oxygen, Wt %	4.7
Distillate (C_4 to 950°F) metals:	
Nickel, ppm-wt	1.0*
Vanadium, ppm-wt	1.0*
Arsenic, ppm-wt	0.5*
Resid (950°F) metals	(NA)
Viscosity @ 100°F, cst	38*

Composition

		Volume %
Butane	C_4	4.9
Naphtha	C_5-400°F	11.6
Distillate	400-950°F	34.5
Bottoms	950°F+	49.0
		100.0

* estimated

(NA) not available

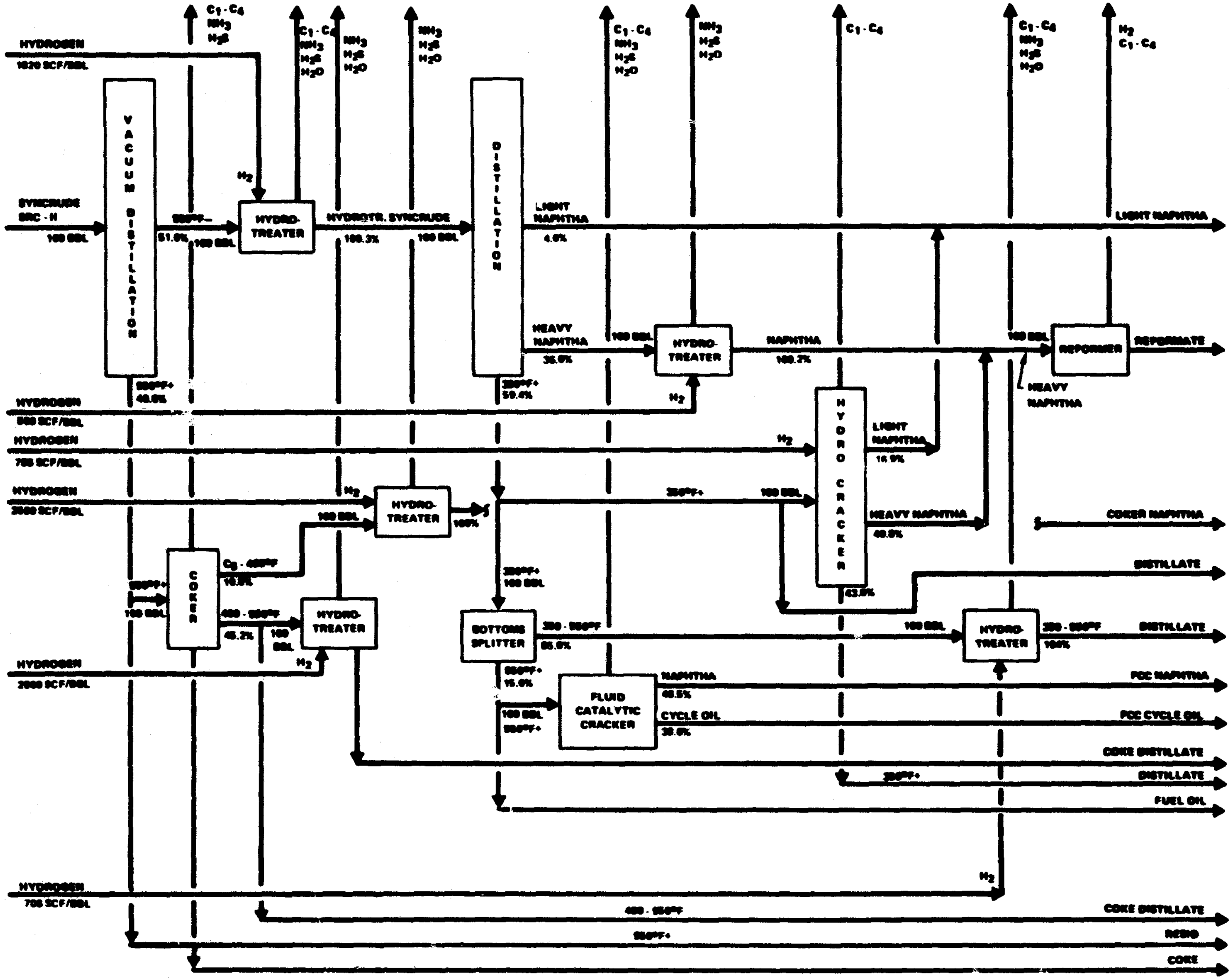

Figure 4-6 - Computer Input Data Diagram, SRC-II Refining, Case 1

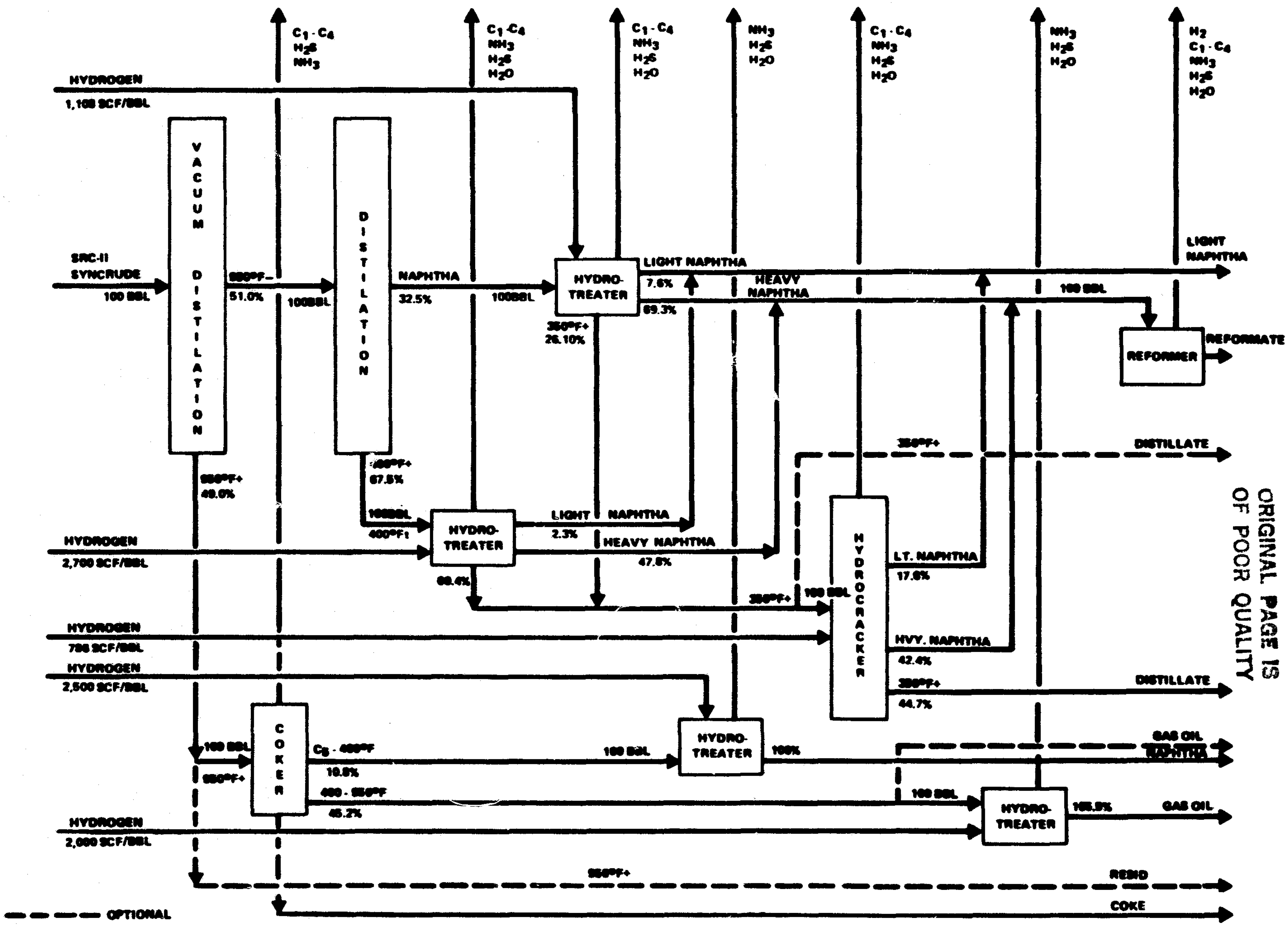

Figure 4-7 - Computer Input Data Diagram,
SRC-II Refining, Case 2

Exhibit 4-A - Linear Programming Computer Run, H-Coal Plus Existing Petroleum Refinery, Case 1

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
	PROFIT	-104.821030				***** OBJECTIVE *****		
****	SLACK VARIABLES							
1	NH3YLD	0.0	-SLACK	0.0	NONE	1.922517	NONE	
2	H2PYLD	0.0	-SLACK	0.0	NONE	2.343104	NONE	
3	H2PLOS	0.0	-SLACK	0.0	NONE	1.081761	NONE	
4	H2UYLD	0.0	-SLACK	0.0	NONE	1.369519	NONE	
5	C1PYLD	0.0	-SLACK	0.0	NONE	4.435727	NONE	
6	C1UYLD	0.0	-SLACK	0.0	NONE	4.543903	NONE	
7	C2-YLD	0.0	-SLACK	0.0	NONE	7.492367	NONE	
8	C2PYLD	0.0	-SLACK	0.0	NONE	7.977982	NONE	
9	C2UYLD	0.0	-SLACK	0.0	NONE	8.086158	NONE	
10	C3-LOS	0.0	-SLACK	0.0	NONE	7.017320	NONE	
11	C3-BAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
12	C3PLOS	0.0	-SLACK	0.0	NONE	7.453234	NONE	
13	C3PBAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
14	C3ULOS	0.0	-SLACK	0.0	NONE	7.287725	NONE	
15	C3UBAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
16	IC4LOS	0.0	-SLACK	0.0	NONE	5.627064	NONE	
17	IC4BAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
18	C4-LOS	0.0	-SLACK	0.0	NONE	4.217804	NONE	
19	C4-BAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
20	NC4LOS	0.0	-SLACK	0.0	NONE	5.042588	NONE	
21	NC4BAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
22	COKYLD	0.0	-SLACK	0.0	NONE	10.000000	NONE	
23	AL3YLD	0.0	-SLACK	0.0	NONE	37.147738	NONE	
24	AL4YLD	0.0	-SLACK	0.0	NONE	34.942260	NONE	
25	S12YLD	0.0	-SLACK	0.0	NONE	0.016418	NONE	
26	S65YLD	0.0	-SLACK	0.0	NONE	0.013917	NONE	
27	S40YLD	0.0	-SLACK	0.0	NONE	0.012240	NONE	
28	S15YLD	0.0	-SLACK	0.0	NONE	0.009361	NONE	
29	S05YLD	0.0	-SLACK	0.0	NONE	0.006953	NONE	
30	S01YLD	0.0	-SLACK	0.0	NONE	0.002448	NONE	
31	CWCYLD	0.0	-SLACK	0.0	NONE	0.161873	NONE	
32	BHPYLD	0.0	-SLACK	0.0	NONE	0.11[illegible]	NONE	
33	KWHYLD	0.0	-SLACK	0.0	NONE	0.138803	NONE	
34	CONYLD	0.0	-SLACK	0.0	NONE	0.000048	NONE	
35	FULYLD	0.0	-SLACK	0.0	NONE	4.998243	NONE	
36	TSWYLD	4787.461221	-SLACK	0.0	NONE	NONE	NONE	MLB TREATED WATER FROM SOUR WATER STR
37	WDWYLD	1744.139012	-SLACK	0.0	NONE	NONE	NONE	MLB WASTE WATER FROM CRUDE DESALTER
38	AWWYLD	463.598060	-SLACK	0.0	NONE	NONE	NONE	MLB ALKYLATION WASTE WATER
39	FULPRD	0.0	-SLACK	0.0	NONE	4.998243	NONE	
40	KWHPRD	0.0	-SLACK	0.0	NONE	0.138303	NONE	
41	TOTNHT	0.0	+SLACK	0.0	NONE	0.374546	NONE	
42	TOTREF	0.0	+SLACK	0.0	NONE	1.904690	NONE	
43	TOTDHT	0.0	+SLACK	0.0	NONE	0.000001	NONE	
44	TOTFCC	0.0	+SLACK	0.0	NONE	0.000001	NONE	
45	TOTREK	0.0	+SLACK	0.0	NONE	0.000001	NONE	
46	TOTGHC	0.0	+SLACK	0.0	NONE	0.000001	NONE	
47	TOTALK	0.0	+SLACK	0.0	NONE	0.000001	NONE	
48	NLGRON	0.0	-SLACK	0.0	NONE	0.106132	NONE	SLACK OF PROPERTIES TO SPECIFICATION
49	NLGRVP	[illegible]	+SLACK	0.0	NONE	0.058429	NONE	" " " " "
50	NO2MXV	[illegible]1782	+SLACK	[illegible]	NONE	NONE	NONE	" " " " "

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	SLACK OF PROPERTIES TO SPECIFICATION
51	NO2MNV	258.758218	-SLACK	0.0	NONE	NONE	NONE	
52	NO2MXS	0.0	+SLACK	0.0	NONE	0.314217	NONE	" " " " "
53	NO2MXD	0.717390	+SLACK	0.0	NONE	NONE	NONE	" " " " "
54	NO6MXV	12.240602	+SLACK	0.0	NONE	NONE	NONE	" " " " "
55	NO6MNV	59.759398	-SLACK	0.0	NONE	NONE	NONE	" " " " "
56	NO6MXS	0.0	+SLACK	0.0	NONE	6.319960	NONE	" " " " "
57	NO6MXD	0.565386	+SLACK	0.0	NONE	NONE	NONE	" " " " "
58	RFOMXV	93.701362	+SLACK	0.0	NONE	NONE	NONE	" " " " "
59	RFOMNV	116.610393	-SLACK	0.0	NONE	NONE	NONE	" " " " -
60	RFOMXS	0.0	+SLACK	0.0	NONE	9.425565	NONE	" " " "
61	RFOMXD	0.393702	+SLACK	0.0	NONE	NONE	NONE	" " " " "
62	TF1MXV	247.805796	+SLACK	0.0	NONE	NONE	NONE	" " " "
63	TF1MNV	72.194204	-SLACK	0.0	NONE	NONE	NONE	" " " "
64	TF1MXS	0.0	+SLACK	0.0	NONE	0.309547	NONE	" " " "
65	TF1MXN	1.177979	+SLACK	0.0	NONE	NONE	NONE	" " " " "
66	TF1MXD	0.650431	+SLACK	0.0	NONE	NONE	NONE	" " " " "
67	TF1X65	0.0	+SLACK	0.0	NONE	6.687595	NONE	" " " " "
68	TF1X95	0.0	+SLACK	0.0	NONE	NONE	NONE	" " " " "
69	LNCYLD	0.0	-SLACK	0.0	NONE	29.189402	NONE	
70	MNCYLD	0.0	-SLACK	0.0	NONE	30.942425	NONE	
71	KECYLD	0.0	-SLACK	0.0	NONE	31.342693	NONE	
72	GOCYLD	0.0	-SLACK	0.0	NONE	31.300408	NONE	
73	RECYLD	0.0	-SLACK	0.0	NONE	30.576705	NONE	
74	GOVYLD	0.0	-SLACK	0.0	NONE	33.020551	NONE	
75	REVYLD	0.0	-SLACK	0.0	NONE	24.241528	NONE	
76	RSTYLD	0.0	-SLACK	0.0	NONE	27.326830	NONE	
77	LHCYLD	0.0	-SLACK	0.0	NONE	30.614364	NONE	
78	HHCYLD	0.0	-SLACK	0.0	NONE	32.365981	NONE	
79	KHCYLD	0.0	-SLACK	0.0	NONE	31.357106	NONE	
80	GHCYLD	0.0	-SLACK	0.0	NONE	31.353533	NONE	
81	GHVYLD	0.0	-SLACK	0.0	NONE	33.668747	NONE	
82	NHKYLD	0.0	-SLACK	0.0	NONE	33.067492	NONE	
83	NAKYLD	0.0	-SLACK	0.0	NONE	31.583426	NONE	
84	GOKYLD	0.0	-SLACK	0.0	NONE	30.964336	NONE	
85	NAFYLD	0.0	-SLACK	0.0	NONE	34.389368	NONE	
86	GAFYLD	0.0	-SLACK	0.0	NONE	32.104466	NONE	
87	DAFYLD	0.0	-SLACK	0.0	NONE	27.011768	NONE	
88	NBFYLD	0.0	-SLACK	0.0	NONE	34.447874	NONE	
89	GBFYLD	0.0	-SLACK	0.0	NONE	32.490673	NONE	
90	DBFYLD	0.0	-SLACK	0.0	NONE	33.065329	NONE	
91	NCFYLD	0.0	-SLACK	0.0	NONE	34.564770	NONE	
92	GCFYLD	0.0	-SLACK	0.0	NONE	31.134101	NONE	
93	DCFYLD	0.0	-SLACK	0.0	NONE	28.211368	NONE	
94	NDFYLD	0.0	-SLACK	0.0	NONE	34.421620	NONE	
95	GDFYLD	0.0	-SLACK	0.0	NONE	31.287897	NONE	
96	DDFYLD	0.0	-SLACK	0.0	NONE	33.980479	NONE	
97	LAHYLD	0.0	-SLACK	0.0	NONE	32.631305	NONE	
98	GHFYLD	0.0	-SLACK	0.0	NONE	32.575163	NONE	
99	HAHYLD	0.0	-SLACK	0.0	NONE	34.729347	NONE	
100	KAHYLD	0.0	-SLACK	0.0	NONE	31.359682	NONE	
101	LBHYLD	0.0	-SLACK	0.0	NONE	32.867942	NONE	
102	HBHYLD	0.0	-SLACK	0.0	NONE	34.715324	NONE	
103	KBHYLD	0.0	-SLACK	0.0	NONE	31.359626	NONE	

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)
104	LCHYLD	0.0	-SLACK	0.0	NONE	32.636388	NONE
105	HCHYLD	0.0	-SLACK	0.0	NONE	34.757392	NONE
106	KCHYLD	0.0	-SLACK	0.0	NONE	33.064329	NONE
107	LDHYLD	0.0	-SLACK	0.0	NONE	33.438242	NONE
108	HDHYLD	0.0	-SLACK	0.0	NONE	34.762067	NONE
109	KDHYLD	0.0	-SLACK	0.0	NONE	31.359852	NONE
110	LKHYLD	0.0	-SLACK	0.0	NONE	32.948318	NONE
111	HKHYLD	0.0	-SLACK	0.0	NONE	32.634048	NONE
112	KKHYLD	0.0	-SLACK	0.0	NONE	31.359186	NONE
113	LVHYLD	0.0	-SLACK	0.0	NONE	32.950188	NONE
114	HVHYLD	0.0	-SLACK	0.0	NONE	32.987759	NONE
115	KVHYLD	0.0	-SLACK	0.0	NONE	31.359023	NONE
116	R5SYLD	0.0	-SLACK	0.0	NONE	35.154902	NONE
117	R1SYLD	0.0	-SLACK	0.0	NONE	35.559131	NONE
118	R5KYLD	0.0	-SLACK	0.0	NONE	35.073861	NONE
119	R1KYLD	0.0	-SLACK	0.0	NONE	35.475286	NONE
120	RK5YLD	0.0	-SLACK	0.0	NONE	35.454047	NONE
121	RK1YLD	0.0	-SLACK	0.0	NONE	35.552412	NONE
122	HNAYLD	0.0	-SLACK	0.0	NONE	33.377396	NONE
123	RHNYLD	0.0	-SLACK	0.0	NONE	35.569940	NONE
124	R5HYLD	0.0	-SLACK	0.0	NONE	35.195393	NONE
125	R1HYLD	0.0	-SLACK	0.0	NONE	35.619021	NONE
126	B1DYLD	0.0	-SLACK	0.0	NONE	31.360336	NONE
127	CH1YLD	0.0	-SLACK	0.0	NONE	29.840496	NONE
128	D1DYLD	0.0	-SLACK	0.0	NONE	31.359968	NONE
129	D1FYLD	0.0	-SLACK	0.0	NONE	32.644292	NONE
130	G1DYLD	0.0	-SLACK	0.0	NONE	31.361492	NONE
131	G1FYLD	0.0	-SLACK	0.0	NONE	34.945207	NONE
132	G1HYLD	0.0	-SLACK	0.0	NONE	31.363055	NONE
133	HHDYLD	0.0	-SLACK	0.0	NONE	32.608066	NONE
134	H1DYLD	0.0	-SLACK	0.0	NONE	30.320932	NONE
135	H1HYLD	0.0	-SLACK	0.0	NONE	34.815821	NONE
136	L1DYLD	0.0	-SLACK	0.0	NONE	33.938920	NONE
137	L1HYLD	0.0	-SLACK	0.0	NONE	34.124023	NONE
138	N1FYLD	0.0	-SLACK	0.0	NONE	34.631964	NONE
139	R11YLD	0.0	-SLACK	0.0	NONE	35.919836	NONE
140	R16YLD	0.0	-SLACK	0.0	NONE	35.289327	NONE

Exhibit 4-A (Cont'd)

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
****	STRUCTURAL VARIABLES							
1	TOTPDV	6697.313760	IN BDS	0.0	NONE	NONE	1.000000	MM$ TOTAL PRODUCT VALUE
2	TOTFDC	6578.512880	IN BDS	0.0	NONE	NONE	-1.000000	MM$ TOTAL FEED COST
3	TOTOPC	26.080300	IN BDS	0.0	NONE	NONE	-1.000000	M$ OPERATING COST (CHEMICALS, WATER)
4	TOTIIC	265.744258	IN BDS	0.0	NONE	NONE	0.0	
5	TOTIC2	87.124052	IN BDS	0.0	NONE	NONE	-2.160000	
6	TOTRCC	21.673976	IN BDS	0.0	NONE	NONE	0.0	MM$ ROYALT. AND CAT. COST OF PET. REFIN.
7	TOTRC2	4.329393	IN BDS	0.0	NONE	NONE	-2.160000	MM$ ROYALT. AND CAT. COST OF NEW REFIN.
8	TOTCAP	287.418234	IN BDS	0.0	NONE	NONE	0.0	
9	ADDCAP	91.453445	IN BDS	0.0	NONE	NONE	0.0	
10	TOTLPG	17.329710	IN BDS	0.0	NONE	NONE	0.0	MBBL TOTAL LPG PRODUCED
11	TOTNLG	108.100000	AT MAX	0.0	108.100000	-6.279682	0.0	MBBL TOTAL NON LEAD GASOLINE PRODUCED
12	MOLNLG	279.819634	IN BDS	0.0	NONE	NONE	0.0	
13	TOTNO2	53.800000	AT MAX	0.0	53.800000	-0.659100	0.0	MBBL TOTAL NO2 FUEL OIL PRODUCED
14	MLBNO2	15.766930	IN BDS	0.0	NONE	NONE	0.0	
15	TOTNO6	8.000000	AT MIN	8.000000	NONE	6.284753	0.0	MBBL TOTAL NO6 FUEL OIL PRODUCED
16	MLBNO6	2.693814	IN BDS	0.0	NONE	NONE	0.0	
17	TOTSUL	0.101424	IN BDS	0.0	NONE	NONE	0.0	MLT TOTAL SULFUR PRODUCED
18	TOTNH3	0.048644	IN BDS	0.0	NONE	NONE	0.0	MT " AMMONIA PRODUCED
19	TOTCOK	0.706131	IN BDS	0.0	NONE	NONE	0.0	MT " COKE FROM COKER PRODUCED
20	TOTRFO	7.010392	IN BDS	0.0	NONE	NONE	0.0	MBBL " REFINERY FUEL OIL
21	MLBRFO	2.462331	IN BDS	0.0	NONE	NONE	0.0	
22	HTVLPG	70.377215	IN BDS	0.0	NONE	NONE	0.0	MMMBTU HEATING VALUE OF PRODUCTS
23	HTVNLG	561.677959	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
24	HTVNO2	304.114379	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
25	HTVNO6	48.902431	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
26	HTVTF1	116.515671	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
27	HTVSUL	0.906490	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
28	HTVNH3	0.940568	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
29	HTVCOK	21.183916	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
30	TOTHTV	1124.618648	IN BDS	0.0	NONE	NONE	0.0	MMMBTU TOTAL HEATING VALUE OF PRODUCTS
31	HTVFED	1245.716275	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " " FEED
32	HTVRFO	44.872224	IN BDS	0.0	NONE	NONE	0.0	MMMBTU HEATING VALUE OF REF. FUEL OIL
33	LHVFUL	77.262981	IN BDS	0.0	NONE	NONE	0.0	MMMBTU LOW HEAT VALUE OF FUEL CONSUMP.
34	TOTTF1	20.000000	AT MIN	20.000000	NONE	31.294244	0.0	MBBL TOTAL TURBINE FUEL PRODUCED
35	MLBTF1	6.105569	IN BDS	0.0	NONE	NONE	0.0	
36	NHTTOT	61.000000	AT MAX	0.0	61.000000	-0.374545	-0.000001	MBBL NAPHTHA HYDROTR. CAPACITY (EXIST.)
37	ADDNHT	3.589743	IN BDS	0.0	NONE	NONE	-0.000002	MBBL ADDITIONAL NAPHTHA H-TR CAPACITY
38	REFTOT	49.000000	AT MAX	0.0	49.000000	-1.904689	-0.000001	MBBL REFORMER CAPACITY (EXIST.)
39	ADDREF	3.067656	IN BDS	0.0	NONE	NONE	-0.000002	MBBL ADDITIONAL REFORMER CAPACITY
40	DHTTOT	2.824896	IN BDS	0.0	22.000000	NONE	-0.000001	MBBL DISTILLATE H-TR. CAPACITY (EXIST.)
41	ADDDHT	0.0	AT MIN	0.0	NONE	1.010665	-0.000002	MBBL ADDIT. DIST. H-TREATER CAPACITY
42	FCCTOT	47.636095	IN BDS	0.0	50.000000	NONE	-0.000001	MBBL FLUID CAT CRACKER CAPACITY (EXIST.)
43	ADDFCC	0.0	AT MIN	0.0	NONE	2.096929	-0.000002	MBBL ADDIT. FCC CAPACITY
44	REKTOT	10.863547	IN BDS	0.0	12.500000	NONE	-0.000001	MBBL RESID COKER CAPACITY (EXIST.)
45	ADDREK	0.0	AT MIN	0.0	NONE	2.473201	-0.000002	MBBL ADDIT. COKER CAPACITY
46	GHCTOT	9.479413	IN BDS	0.0	10.300000	NONE	-0.000001	MBBL GAS OIL H-CRACKER CAPACITY (EXIST.)
47	ADDGHC	0.0	AT MIN	0.0	NONE	5.750785	-0.000002	MBBL ADDIT. H-CRACKER CAPACITY
48	ALKTOT	3.090654	IN BDS	0.0	8.000000	NONE	-0.000001	MBBL ALKYLATION CAPACITY (EXIST.)
49	ADDALK	0.0	AT MIN	0.0	NONE	2.452465	-0.000002	MBBL ADDIT. ALKYLATION CAPACITY
50	H2PPLT	0.0	AT MAX	0.0	0.0	-0.486650	0.0	MMMSCF HYDROGEN PLANT CAPACITY (EXIST.)
51	ADDH2P	3.663364	IN BDS	0.0	NONE	NONE	-0.000002	MMMSCF ADDIT. HYDROGEN PLANT CAPACITY

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WTS

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
52	POXPLT	0.0	AT MIN	0.0	NONE	3.821319	0.0	MMSCF PARTIAL OXIDATION PLANT CAPACITY
53	COKPUR	0.0	AT MIN	0.0	NONE	5.000000	0.0	
54	H2PPUR	0.0	AT MIN	0.0	NONE	7.656896	0.0	
55	SULPLT	0.101424	IN BDS	0.0	0.135000	NONE	0.0	MLT SULFUR PLANT CAPACITY (EXIST.)
56	ADDSUL	0.0	AT MIN	0.0	NONE	136.753058	-0.000002	MLT ADDIT. SULFUR PLANT
57	SWSPLT	5300.000000	AT MAX	0.0	5300.000000	-0.000890	0.0	MLB SOUR WATER STRIPPER CAPACITY (EXIST
58	ADDSWS	959.386334	IN BDS	0.0	NONE	NONE	-0.000002	MLB ADDIT. SOUR WATER STRIPPER CAPACITY
59	CONTSW	0.0	AT MIN	0.0	NONE	0.000048	0.0	
60	NH3PLT	0.017000	AT MAX	0.0	0.017000	-61.482242	0.0	MT AMMONIA PLANT CAPACITY (EXIST.)
61	ADDNH3	0.031644	IN BDS	0.0	NONE	NONE	-0.000002	MT ADDIT. AMMONIA PLANT CAPACITY
62	CWCPLT	185.696640	IN BDS	0.0	196.000000	NONE	0.0	MGAL COOLING WATER (CIRC) CAPAC. (EXIST
63	ADDCWC	0.0	AT MIN	0.0	NONE	0.060482	-0.000002	MGAL ADDIT. COOL. WATER CAPACITY
64	CONPLT	44566.429897	IN BDS	0.0	NONE	NONE	0.0	
65	KWHPLT	1277.730211	IN BDS	0.0	NONE	NONE	0.0	MKWH POWER PLANT POWER PRODUCTION
66	S12PLT	15100.000000	AT MAX	0.0	15100.000000	-0.006912	0.0	MLB POWER PLT 1250 PSI STEAM PROD. (EXI
67	ADDS12	444.323764	IN BDS	0.0	NONE	NONE	-0.000000	MLB ADDIT. POWER PLANT STEAM PRODUCTION
68	KWHPT1	281.141685	IN BDS	0.0	NONE	NONE	0.0	
69	KWHPT2	179.541759	IN BDS	0.0	NONE	NONE	0.0	
70	KWHPT3	282.823073	IN BDS	0.0	NONE	NONE	0.0	
71	KWHPT4	149.629517	IN BDS	0.0	NONE	NONE	0.0	
72	KWHPT5	214.608239	IN BDS	0.0	NONE	NONE	0.0	
73	KWHPT6	169.985937	IN BDS	0.0	NONE	NONE	0.0	
74	TOTCOF	1.097159	IN BDS	0.0	NONE	NONE	0.0	
75	TOTCWD	0.0	IN BDS	0.0	NONE	NONE	0.0	
76	TOTBHP	241.405632	IN BDS	0.0	NONE	NONE	0.0	
77	BHPMOT	241.405632	IN BDS	0.0	NONE	NONE	0.0	
78	BHPTUR	0.0	IN BDS	0.0	NONE	NONE	0.0	
79	TOTBTU	22.434428	IN BDS	0.0	NONE	NONE	0.0	
80	C3-ALK	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL C3- UNSAT TO ALKYLATION
81	C4-ALK	1.776236	IN BDS	0.0	NONE	NONE	0.0	MBBL C4- UNSAT TO ALKYLATION
82	C1PHPT	0.915841	IN BDS	0.0	NONE	NONE	0.0	MMSCF C1 TO HYDROGEN PLANT
83	C2PHPT	0.0	AT MIN	0.0	NONE	0.215460	0.0	MMSCF C2 " " "
84	C3PHPT	0.0	AT MIN	0.0	NONE	6.726641	0.0	MBBL C3 " " "
85	H2PFUL	0.914565	IN BDS	0.0	NONE	NONE	0.0	MMSCF LIGHT GASES TO FUEL
86	C1PFUL	2.915542	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
87	C2PFUL	3.974996	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
88	C3PFUL	0.523264	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
89	IC4FUL	0.254369	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
90	NC4FUL	0.159001	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
91	H2UFUL	2.238914	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
92	C1UFUL	7.003638	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
93	C2-FUL	2.185572	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
94	C2UFUL	3.766843	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
95	C3-FUL	0.659970	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
96	C3UFUL	0.345831	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
97	C4-FUL	0.277946	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
98	TOTC3-	4.299852	IN BDS	0.0	NONE	NONE	0.0	
99	TOTC3P	3.468426	IN BDS	0.0	NONE	NONE	0.0	
100	TOTC3U	2.305540	IN BDS	0.0	NONE	NONE	0.0	
101	TOTIC4	2.826542	IN BDS	0.0	NONE	NONE	0.0	
102	TOTI4U	3.535587	IN BDS	0.0	NONE	NONE	0.0	
103	TOTC4-	4.632471	IN BDS	0.0	NONE	NONE	0.0	
104	TOTNC4	7.[illegible]0028	IN BDS	0.0	NONE	NONE	0.0	

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Exhibit 4-A (Cont'd)

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
105	TOTN6U	1.057096	IN BDS	0.0	NONE	NONE	0.0	
106	TOTSFG	9.145646	IN BDS	0.0	NONE	NONE	0.0	
107	TOTH2H	76.862418	IN BDS	0.0	NONE	NONE	0.0	
108	CRDCRU	165.950429	IN BDS	0.0	200.000000	NONE	0.0	MBBL PETROLEUM CRUDE TO CRUDE DIST.
109	LNCLHT	12.645423	IN BDS	0.0	NONE	NONE	0.0	MBBL LT NAPHTHA FROM CRUDE UNIT TO H-TR.
110	HNCNHT	34.235574	IN BDS	0.0	NONE	NONE	0.0	MBBL HVY " " " " " "
111	KECKHT	0.0	AT MIN	0.0	NONE	1.024229	0.0	MBBL KEROSENE " " " " " "
112	GOCGHT	0.0	AT MIN	0.0	NONE	0.977220	0.0	MBBL ATM GAS OIL " " " " "
113	GOCHCR	5.742353	IN BDS	0.0	NONE	NONE	0.0	MBBL ATM GAS OIL TO HYDROCRACKER
114	REGVAC	62.148436	IN BDS	0.0	NONE	NONE	0.0	MBBL RESID FROM CRUDE TO VACUUM UNIT
115	GOVHCR	0.0	AT MIN	0.0	NONE	0.933348	0.0	MBBL VACUUM GAS OIL TO HYDROCRACKER
116	GOVFC7	0.0	AT MIN	0.0	NONE	0.515697	0.0	MBBL " " " " FCC 75% CONVR.
117	GOVFC8	45.293760	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " FCC 85% "
118	GOVGVH	0.0	AT MIN	0.0	NONE	1.492965	0.0	IBTH " " " " HYDROTREATER
119	GHVFC7	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. VAC GAS OIL TO FCC 75% CONVR.
120	GHVFC8	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " " 85% "
121	RFVDCK	10.863547	IN BDS	0.0	NONE	NONE	0.0	MBBL VACUUM RESID (LT ARAB) TO COKER
122	RSTDCK	0.0	AT MIN	0.0	NONE	1.109292	0.0	MBBL " " (SO. TEX) " "
123	NAKNHT	3.204746	IN BDS	0.0	NONE	NONE	0.0	MBBL COKER NAPHTHA TO HYDROTREATER
124	GOKHCR	3.737060	IN BDS	0.0	NONE	NONE	0.0	MBBL COKER GAS OIL TO HYDROCRACKER
125	GAFHTR	0.0	AT MIN	0.0	NONE	0.970364	0.0	MBBL FCC LC-GAS OIL TO H-TREATER
126	GAFHCR	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " HYDROCRACKER
127	GCFHTR	2.624696	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " H-TREATER
128	GCFHCR	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " HYDROCRACKER
129	GBFHCR	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
130	GDFHCR	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
131	GHFFUO	2.824896	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. FCC LC GAS OIL TO FUEL OIL
132	HHCR95	0.0	AT MIN	0.0	NONE	0.070653	0.0	MBBL H-TR. SR NAPHTHA TO 95 RON REFORMER
133	HHCR10	34.235574	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " 100 " "
134	NHKR95	3.204746	IN BDS	0.0	NONE	NONE	0.0	MBBL " COKER NAPH " 95 " "
135	NHKR10	0.0	AT MIN	0.0	NONE	0.016184	0.0	MBBL " " " " 100 " "
136	HAHHNR	0.0	AT MIN	0.0	NONE	1.351951	0.0	MBBL HYDROCRACKER NAPHTHA TO REFORMER
137	HBHHNR	0.0	AT MIN	0.0	NONE	1.337928	0.0	MBBL " " " "
138	HCHHNR	0.0	AT MIN	0.0	NONE	1.379997	0.0	MBBL " " " "
139	HDHHNR	0.0	AT MIN	0.0	NONE	1.384671	0.0	MBBL " " " "
140	HNAR10	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
141	HKHR95	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
142	HKHR10	2.914907	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
143	HVHR95	0.0	AT MIN	0.0	NONE	1.367241	0.0	MBBL " " " "
144	HVHR10	4.180433	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
145	LHCHPT	0.0	AT MIN	0.0	NONE	8.990195	0.0	MBBL H-TR LT. NAPHTHA SR TO HYDROGEN PLT
146	CO1HTR	50.000000	FIXED	50.000000	50.000000	7.385247	0.0	MBBL RAW H-COAL OIL TO HYDROTREATER
147	CH1DIS	51.800000	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. H-COAL OIL TO DISTILLATION
148	H1DHTR	14.504000	IN BDS	0.0	NONE	NONE	0.0	MBBL HVY NAPHTHA FROM H-COAL-DIST TO H-TR.
149	B1DHCR	0.0	AT MIN	0.0	NONE	5.066363	0.0	MBBL H-COAL DIST. BOTTOMS TO H-CRACKER
150	B1DDIS	29.267000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " SPLITTER
151	G1DFCC	2.542315	IN BDS	0.0	NONE	NONE	0.0	MBBL HVY GAS OIL FROM SPLITTER TO FCC
152	NAHREF	0.0	AT MIN	0.0	NONE	2.207756	0.0	MBBL H-TR. NAPHTHA TO REFORMER
153	HHDR96	7.531996	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. NAPHTHA TO 96 RON REFORMER
154	HHDR10	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " 100 " "
155	L1DHPT	0.0	AT MIN	0.0	NONE	12.314751	0.0	MBBL LT. NAPHTHA TO HYDROGEN PLANT
156	L1HHPT	0.0	AT MIN	0.0	NONE	12.499854	0.0	MBBL " " " " "
157	C3-LPG	3.439882	IN BDS	0.0	NONE	NONE	0.0	MBBL C3 & C4 TO LPG

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H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
158	C3PLPG	2.965162	IN BDS	0.0	NONE	NONE	0.0	MBBL C3 & C4 TO LPG
159	C3ULPG	1.959709	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
160	C4-LPG	2.578285	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
161	IC4LPG	4.047304	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
162	NC4LPG	2.339367	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
163	IC4NLG	0.0	AT MIN	0.0	NONE	1.203460	0.0	MBBL C4 TO GASOLINE
164	NC4NLG	6.508757	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
165	C4-NLG	0.0	AT MIN	0.0	NONE	1.770925	0.0	MBBL " " "
166	AL3NLG	0.0	AT MIN	0.0	NONE	2.529674	0.0	MBBL ALKYLATE TO GASOLINE
167	AL4NLG	3.090654	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
168	LHCNLG	12.645423	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. LT. NAPHTHA SR TO GASOLINE
169	LAHNLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL H-CRACK LT. NAPHTHA " "
170	LCHNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
171	LBHNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
172	LDHNLG	0.0	AT MIN	0.0	NONE	0.613247	0.0	MBBL " " " " "
173	LKHNLG	1.046377	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
174	LVHNLG	1.665282	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
175	MAFNLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL FCC-NAPHTHA TO GASOLINE
176	NCFNLG	27.629206	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
177	NBFNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
178	NDFNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
179	HAHNLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL HVY H-CRACKER NAPHTHA TO GASOLINE
180	HCHNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
181	HBHNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
182	HDHNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
183	HKHNLG	0.0	AT MIN	0.0	NONE	0.423292	0.0	MBBL " " " " "
184	R5SNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL REFORMATE TO GASOLINE
185	R1SNLG	28.107406	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
186	R5KNLG	2.587833	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
187	R1KNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
188	RHNNLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
189	RK5NLG	0.0	AT MIN	0.0	NONE	0.299145	0.0	MBBL " " "
190	RK1NLG	2.503905	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
191	R5HNLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
192	R1HNLG	3.260738	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
193	KECNO2	26.5[illegible]259	IN BDS	0.0	NONE	NONE	0.0	MBBL KEROSENE SR TO NO2 FUEL OIL
194	KHCNO2	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL HT KEROSENE SR TO NO2 FUEL OIL
195	KAHNO2	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL H-CRACK. KEROSENE TO NO2 FUEL OIL
196	KCHNO2	0.0	AT MIN	0.0	NONE	1.704421	0.0	MBBL " " " " " "
197	KBHNO2	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
198	KDHNO2	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
199	KKHNO2	0.056056	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
200	KVHNO2	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
201	GAFNO2	0.0	AT MIN	0.0	NONE	0.931139	0.0	MBBL FCC LC-GAS OIL TO NO2 FUEL OIL
202	GCFNO2	0.0	AT MIN	0.0	NONE	0.009956	0.0	MBBL " " " " " " "
203	GBFNO2	0.0	AT MIN	0.0	NONE	1.188514	0.0	MBBL " " " " " " "
204	GDFNO2	0.0	AT MIN	0.0	NONE	0.007164	0.0	MBBL " " " " " " "
205	GOCNO2	9.535482	IN BDS	0.0	NONE	NONE	0.0	MBBL ATM GAS OIL SR TO NO2 FUEL OIL
206	GHCNO2	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " H-TR TO NO2 FO
207	KECNO6	0.0	AT MIN	0.0	NONE	0.566337	0.0	MBBL KEROSENE SR TO NO6 FUEL OIL
208	KHCNO6	0.0	AT MIN	0.0	NONE	0.312907	0.0	MBBL HT KEROSENE SR TO NO6 FO
209	GAFNO6	0.0	AT MIN	0.0	NONE	4.504866	0.0	MBBL FCC LC GAS OIL TO NO6 FUEL OIL
210	GCFNO6	0.0	AT MIN	0.0	NONE	4.465053	0.0	MBBL " " " " " " " "

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
211	GBFN06	0.0	AT MIN	0.0	NONE	2.344825	0.0	MBBL FCC LC-GAS OIL TO NO6 FUEL OIL
212	GDFN06	0.0	AT MIN	0.0	NONE	1.484527	0.0	MBBL " " " " " "
213	GHFN06	0.0	AT MIN	0.0	NONE	1.576437	0.0	MBBL HT FCC LC-GAS OIL TO NO6 FUEL OIL
214	GOCN06	0.0	AT MIN	0.0	NONE	1.325768	0.0	MBBL ATM GAS OIL SR TO NO6 FUEL OIL
215	GHCN06	0.0	AT MIN	0.0	NONE	0.318972	0.0	MBBL " " " " H-TR TO NO6 FUEL OIL
216	GOKN06	0.0	AT MIN	0.0	NONE	3.330672	0.0	MBBL COKER GAS OIL TO NO6 FUEL OIL
217	GOVN06	0.0	AT MIN	0.0	NONE	4.574456	0.0	MBBL VAC. GAS OIL TO NO6 FUEL OIL
218	GHVN06	0.0	AT MIN	0.0	NONE	3.200365	0.0	MBBL H-TR VAC GAS OIL TO NO6 FUEL OIL
219	REVN06	0.0	AT MIN	0.0	NONE	1.734300	0.0	MBBL VACUUM RESID (LT ARAB) TO NO6 FO
220	RSTN06	4.117794	IN BDS	0.0	NONE	NONE	0.0	MBBL " " (SO. TEX) " " FO
221	DAFN06	0.0	AT MIN	0.0	NONE	0.819498	0.0	MBBL FCC DECANT OIL TO NO6 FUEL OIL
222	DCFN06	0.0	AT MIN	0.0	NONE	1.342104	0.0	MBBL " " " " " " "
223	DBFN06	0.0	AT MIN	0.0	NONE	2.686717	0.0	MBBL " " " " " " "
224	DDFN06	0.0	AT MIN	0.0	NONE	3.644337	0.0	MBBL " " " " " " "
225	KECRFO	0.0	AT MIN	0.0	NONE	2.457580	0.0	MBBL KEROSENE SR TO REFINERY FUEL
226	KHCRFO	0.0	AT MIN	0.0	NONE	2.051965	0.0	MBBL HT KEROSENE SR TO REFINERY FUEL
227	GAFRFO	0.0	AT MIN	0.0	NONE	5.275860	0.0	MBBL FCC LC-GAS OIL TO REF. FUEL
228	GCFRFO	0.0	AT MIN	0.0	NONE	5.147258	0.0	MBBL " " " " " "
229	GBFRFO	0.0	AT MIN	0.0	NONE	2.414933	0.0	MBBL " " " " " "
230	GDFRFO	0.0	AT MIN	0.0	NONE	0.756643	0.0	MBBL " " " " " "
231	GHFRFO	2.824896	IN BDS	0.0	NONE	NONE	0.0	MBBL HT FCC LC GAS OIL TO REF. FUEL
232	GOCRFO	0.0	AT MIN	0.0	NONE	2.626255	0.0	MBBL ATM GAS OIL SR TO REF. FUEL
233	GHCRFO	0.0	AT MIN	0.0	NONE	1.144202	0.0	MBBL " " " " H-TR TO REF. FUEL
234	GOKRFO	0.0	AT MIN	0.0	NONE	4.499931	0.0	MBBL COKER GAS OIL TO REF. FUEL
235	GOVRFO	0.0	AT MIN	0.0	NONE	5.382476	0.0	MBBL VAC GAS OIL TO REF. FUEL
236	GHVRFO	0.0	AT MIN	0.0	NONE	3.145706	0.0	MBBL HT VAC GAS OIL TO REF. FUEL
237	REVRFO	0.0	AT MIN	0.0	NONE	3.105485	0.0	MBBL VAC RESID (LT ARAB) TO REF. FUEL
238	RSTRFO	1.873315	IN BDS	0.0	NONE	NONE	0.0	MBBL " " (SO. TEX) " "
239	DAFRFO	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL FCC DECANT OIL TO REF. FUEL
240	DCFRFO	1.358613	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
241	DBFRFO	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " "
242	DDFRFO	0.0	AT MIN	0.0	NONE	0.656452	0.0	MBBL " " " " " "
243	L1DNLG	3.988600	IN BDS	0.0	NONE	NONE	0.0	MBBL H-COAL LT-NAPHTHA TO GASOLINE
244	L1HNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL H-COAL HYDROCR. LT NAPHT. TO GASOL.
245	N1FNLG	1.164380	IN BDS	0.0	NONE	NONE	0.0	MBBL " FCC NAPHTHA TO GASOLINE
246	H1HNLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " HYDROCR. HVY NAPHT. TO GASOL.
247	HHDNLG	6.972004	IN BDS	0.0	NONE	NONE	0.0	MBBL " HYDROTR. HVY NAPHT. " "
248	R16NLG	6.929436	IN BDS	0.0	NONE	NONE	0.0	MBBL " 96 RON REFORMATE " "
249	R11NLG	0.0	AT MIN	0.0	NONE	0.256409	0.0	MBBL " 100 " " " "
250	D1DN02	16.994377	IN BDS	0.0	NONE	NONE	0.0	MBBL " DISTILLATE TO NO2 FUEL OIL
251	G1DN02	0.628626	IN BDS	0.0	NONE	NONE	0.0	MBBL " GAS OIL TO NO2 FUEL OIL
252	G1HN02	0.0	AT MIN	0.0	NONE	0.002800	0.0	MBBL " H-CRACK. GAS OIL TO NO2 FO
253	D1FN02	0.0	AT MIN	0.0	NONE	1.483359	0.0	MBBL " FCC LC GAS OIL TO NO2 FO
254	D1DN06	0.0	AT MIN	0.0	NONE	0.157107	0.0	MBBL " DISTILLATE TO NO6 FUEL OIL
255	G1DN06	3.882206	IN BDS	0.0	NONE	NONE	0.0	MBBL " GAS OIL TO NO6 FUEL OIL
256	G1HN06	0.0	AT MIN	0.0	NONE	0.083090	0.0	MBBL " H-CRACK. GAS OIL TO NO6 FO
257	D1FN06	0.0	AT MIN	0.0	NONE	1.372200	0.0	MBBL " FCC LC GAS OIL TO NO6 FO
258	G1FN06	0.0	AT MIN	0.0	NONE	3.242437	0.0	MBBL " FCC HVY GAS OIL TO NO6 FO
259	D1DRFO	0.0	AT MIN	0.0	NONE	1.908655	0.0	MBBL " DISTILLATE TO REFINERY FUEL
260	G1DRFO	0.0	AT MIN	0.0	NONE	0.533997	0.0	MBBL " GAS OIL TO REF. FUEL
261	G1HRFO	0.0	AT MIN	0.0	NONE	0.514699	0.0	MBBL " H-CRACK. GAS OIL TO REF. FUEL
262	D1FRFO	0.854218	IN BDS	0.0	NONE	NONE	0.0	MBBL " FCC LC GAS OIL TO REF. FUEL
263	G1FRFO	0.099150	IN BDS	0.0	NONE	NONE	0.0	MBBL " FCC HVY GAS OIL TO REF. FUEL

H-COAL CASE 1 PLUS PETROLEUM
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
264	KECTF1	0.0	AT MIN	0.0	NONE	0.002733	0.0	MBBL KEROSENE SR HVY GO TO TURBINE FUEL
265	KHCTF1	0.0	AT MIN	0.0	NONE	0.003010	0.0	MBBL HT KEROSENE SR " " "
266	KAHTF1	0.0	AT MIN	0.0	NONE	0.000649	0.0	MBBL H-CRACK. KEROSENE " " "
267	KCHTF1	0.0	AT MIN	0.0	NONE	1.704516	0.0	MBBL " " " " "
268	KBHTF1	0.0	AT MIN	0.0	NONE	0.000787	0.0	MBBL " " " " "
269	KDHTF1	0.0	AT MIN	0.0	NONE	0.000234	0.0	MBBL " " " " "
270	KKHTF1	0.0	AT MIN	0.0	NONE	0.001864	0.0	MBBL " " " " "
271	KVHTF1	0.0	AT MIN	0.0	NONE	0.002264	0.0	MBBL " " " " "
272	GAFTF1	0.0	AT MIN	0.0	NONE	0.923709	0.0	MBBL FCC LC GAS OIL " " "
273	GCFTF1	2.610358	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " " "
274	GBFTF1	0.0	AT MIN	0.0	NONE	1.184377	0.0	MBBL " " " " " " "
275	GDFTF1	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " " " "
276	GHFTF1	0.0	AT MIN	0.0	NONE	1.228450	0.0	MBBL HT FCC LT GAS OIL " " "
277	GOCTF1	12.170365	IN BDS	0.0	NONE	NONE	0.0	MBBL ATM GAS OIL SR " " "
278	GHCTF1	0.0	AT MIN	0.0	NONE	0.001053	0.0	MBBL HT. ATM GAS OIL SR " " "
279	GOKTF1	0.0	AT MIN	0.0	NONE	3.124340	0.0	MBBL COKER GAS OIL " " "
280	GOVTF1	0.0	AT MIN	0.0	NONE	8.484341	0.0	MBBL VACUUM GAS OIL " " "
281	GHVTF1	0.0	AT MIN	0.0	NONE	9.223366	0.0	MBBL HT VAC GAS OIL " " "
282	REVTF1	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL VAC RESID (LT ARAB) " " "
283	RSTTF1	0.0	AT MIN	0.0	NONE	2.648041	0.0	MBBL VAC RESID (SO. TEX) " " "
284	DAFTF1	0.0	AT MIN	0.0	NONE	2.591539	0.0	MBBL FCC DECANT OIL " " "
285	DCFTF1	0.0	AT MIN	0.0	NONE	3.758519	0.0	MBBL " " " " " "
286	DBFTF1	0.0	AT MIN	0.0	NONE	8.439277	0.0	MBBL " " " " " "
287	DDFTF1	0.0	AT MIN	0.0	NONE	9.357444	0.0	MBBL " " " " " "
288	D1DTF1	5.219276	IN BDS	0.0	NONE	NONE	0.0	MBBL H-COAL DISTILLATE " " "
289	G1DTF1	0.0	AT MIN	0.0	NONE	3.339861	0.0	MBBL " GAS OIL " " "
290	G1HTF1	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " H-CR. GAS OIL " " "
291	D1FTF1	0.0	AT MIN	0.0	NONE	1.475684	0.0	MBBL " FCC LC GO " " "
292	G1FTF1	0.0	AT MIN	0.0	NONE	10.256750	0.0	MBBL " FCC HVY GO " " "

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SECTION 5

NEW REFINERIES TO UPGRADE FUELS

When coal and oil shale derived liquid crudes become available in sizeable quantities it is anticipated that for economic, and possibly technical reasons, it may be advantageous to design and build new refineries specific to processing and upgrading the synthetic crudes. Accordingly, the purpose of this section is to present the design, operation and economics for new refineries, specific to the upgrading of the syncrude process products.

These new refineries are equivalent to the equipment additions made to the existing typical petroleum refinery as covered in Section 4, with the exception that hydrogen facilities are added. These are individual stand-alone refineries with the attendant services and offsites. As was done in Section 4, computer models for the individual liquid to be processed were developed and used to optimize processing and economics.

A copy of representative results from an individual computer run is included as Exhibit 5-A at the end of Section 5.

5.1 SHALE OIL REFINERY

The new refinery to upgrade shale oil is based on the Figure 4-2 and 4-3, Section 4, diagrams used to depict the shale portion of the combined refinery in Section 4. Since this is a stand-alone refinery, it will require an additional source of hydrogen because the reformer hydrogen source is not adequate.

5.1.1 ADDITIONS TO SHALE OIL MODEL INPUT DATA

A. Hydrogen Plant Addition

The hydrogen plant addition to Figures 4-2 and 4-3, Section 4, is based on the partial oxidation of petroleum coke to hydrogen using process conversion units as follows:

- Gasifier and Quench Unit
- Oxygen Plant
- Acid Gas Removal Unit
- Shift Conversion Unit
- CO_2 Removal Unit
- Methanator Unit

B. Feed and Product Analysis

The petroleum coke feed elemental analysis to the partial oxidation unit is as follows:

Element	wt%
Carbon	90.8
Hydrogen	3.3
Nitrogen	0.8
Oxygen	3.1
Sulfur	0.8
Ash	1.2
	100.0

The hydrogen product from the partial oxidation and shift routes contains about 97% (vol) hydrogen. The remaining major constituents are methane and inerts.

C. Investment Cost Data

Investment cost data for the partial oxidation of coke to hydrogen complex was based on in-house estimates for a 50 million SCFD hydrogen plant processing petroleum coke. The reference date for all cost data is March 31, 1980. Capacity ratio exponents (powers) were based on past experience with coal conversion process unit costs.

D. Operating Cost Data

Operating cost is based on catalyst and chemical usage. Chemical usage is based on in-house and licensor data on refining units. Catalyst costs and usage are based on in-house refining units. Chemical costs were taken from "Chemical Marketing Reporter" publication.

5.2 H-COAL REFINERY

The new refinery to upgrade H-Coal is based on the Figures 4-4 and 4-5 in Section 4.3. These diagrams depict the H-Coal portion of the combined refining discussed in Section 4.3.

5.2.1 ADDITIONS TO H-COAL MODEL INPUT DATA

A. Hydrogen Plant Addition.

The stand-alone refinery to upgrade H-Coal liquids produces adequate hydrogen from the conversion of refinery gases and the reformer byproduct hydrogen. An additional source of hydrogen is not required.

5.3 SRC-II REFINERY

The new refinery to upgrade SRC-II oil is based on the Figure 4-6 and 4-7, Section 4 diagrams used to depict the SRC-II portion of the combined refinery in Section 4.2. Since this is a stand-alone refinery, it will require an additional source of hydrogen because the conversion of refinery light gases and the reformer hydrogen source in a new shale oil facility are not adequate.

5.3.1 ADDITIONS TO SRC-II OIL MODEL INPUT DATA

A. Hydrogen Plant Addition

The hydrogen plant addition to Figures 4-6 and 4-7, Section 4, is based on the partial oxidation of SRC-II 950°F plus fraction to hydrogen using process conversion units as follows:

- o Gasifier and Quench Unit
- o Oxygen Plant
- o Acid Gas Removal Unit
- o Shift Conversion Unit
- o CO_2 Removal Unit
- o Methanator Unit

B. Feed and Product Analysis

The SRC-II 950°F plus fraction, part of which is feed to the partial oxidation unit, is a heavy coal tar which must be pumped hot. It will contain some ash and unconverted coal. The sulfur content is about 0.5 vol % and the gravity is about -8.9 °API. The hydrogen product from the partial oxidation and shift routes contains about 97% (vol) hydrogen. The remaining major constituents are methane and inerts.

C. Investment Cost Data

Investment cost data for the partial oxidation of SRC-II resid in the hydrogen complex is based on in-house estimates for a 50 million SCFD hydrogen plant. The reference date for all cost data is March 1980. Capacity ratio exponentials are based on past experience with refinery process unit costs.

D. Operating Cost Data

Operating cost is based on catalyst and chemical usage. Chemical usage is based on in-house and licensor data for refining units. Catalyst costs and usage are based on in-house data for refining units. Chemical costs were taken from "Chemical Marketing Reporter" publication.

Exhibit 5-A - Linear Programming Computer Run, New H-Coal Oil Refinery, Case 1

H-COAL CASE 1
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
	PROFIT	-505.004799				* * * * * OBJECTIVE * * * * *		
****	SLACK VARIABLES							
1	NH3YLD	0.0	-SLACK	0.0	NONE	1.803305	NONE	
2	H2PYLD	0.0	-SLACK	0.0	NONE	3.145540	NONE	
3	H2PLOS	0.0	-SLACK	0.0	NONE	2.260878	NONE	
4	H2UYLD	0.0	-SLACK	0.0	NONE	1.110750	NONE	
5	C1PYLD	0.0	-SLACK	0.0	NONE	3.459251	NONE	
6	C1UYLD	0.0	-SLACK	0.0	NONE	3.685339	NONE	
7	C2-YLD	0.0	-SLACK	0.0	NONE	6.076695	NONE	
8	C2PYLD	0.0	-SLACK	0.0	NONE	6.332202	NONE	
9	C2UYLD	0.0	-SLACK	0.0	NONE	6.558290	NONE	
10	C3-LOS	0.0	-SLACK	0.0	NONE	NONE	NONE	
11	C3-BAL	0.0	-SLACK	0.0	NONE	28.821120	NONE	
12	C3PLOS	0.0	-SLACK	0.0	NONE	10.980349	NONE	
13	C3PBAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
14	C3ULOS	0.0	-SLACK	0.0	NONE	NONE	NONE	
15	C3UBAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
16	IC4LOS	0.0	-SLACK	0.0	NONE	9.627891	NONE	
17	IC4BAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
18	C4-LOS	0.0	-SLACK	0.0	NONE	NONE	NONE	
19	C4-BAL	0.0	-SLACK	0.0	NONE	33.032460	NONE	
20	NC4LOS	0.0	-SLACK	0.0	NONE	8.998351	NONE	
21	NC4BAL	0.0	-SLACK	0.0	NONE	25.000000	NONE	
22	COKYLD	0.0	-SLACK	0.0	NONE	10.000000	NONE	
23	AL3YLD	0.0	-SLACK	0.0	NONE	41.293325	NONE	
24	AL4YLD	0.0	-SLACK	0.0	NONE	41.533869	NONE	
25	S12YLD	0.0	-SLACK	0.0	NONE	0.014836	NONE	
26	S65YLD	0.0	-SLACK	0.0	NONE	0.012472	NONE	
27	S40YLD	0.0	-SLACK	0.0	NONE	0.010888	NONE	
28	S15YLD	0.0	-SLACK	0.0	NONE	0.008167	NONE	
29	S05YLD	0.0	-SLACK	0.0	NONE	0.005892	NONE	
30	S01YLD	0.0	-SLACK	0.0	NONE	0.001634	NONE	
31	CWCYLD	0.0	-SLACK	0.0	NONE	0.214161	NONE	
32	BHPYLD	0.0	-SLACK	0.0	NONE	0.108683	NONE	
33	KWHYLD	0.0	-SLACK	0.0	NONE	0.131196	NONE	
34	CONYLD	0.0	-SLACK	0.0	NONE	0.000048	NONE	
35	FULYLD	0.0	-SLACK	0.0	NONE	4.053832	NONE	
36	TSWYLD	2415.682412	-SLACK	0.0	NONE	NONE	NONE	MLB TREATED WATER FROM SOUR WATER STR.
37	WDWYLD	0.0	-SLACK	0.0	NONE	NONE	NONE	
38	AWWYLD	0.000000	-SLACK	0.0	NONE	NONE	NONE	
39	KWHPRD	0.0	-SLACK	0.0	NONE	0.130696	NONE	
40	TOTNHT	0.0	+SLACK	0.0	NONE	0.000001	NONE	
41	TOTREF	0.0	+SLACK	0.0	NONE	0.000001	NONE	
42	TOTDHT	0.0	+SLACK	0.0	NONE	NONE	NONE	
43	TOTFCC	0.0	+SLACK	0.0	NONE	NONE	NONE	
44	TOTREK	0.0	+SLACK	0.0	NONE	NONE	NONE	
45	TOTGHC	0.0	+SLACK	0.0	NONE	0.000001	NONE	
46	TOTALK	0.000000	+SLACK	0.0	NONE	NONE	NONE	
47	NLGRON	0.0	-SLACK	0.0	NONE	0.080181	NONE	SLACK OF PROPERTIES TO SPECIFICATION
48	NLGRVP	0.0	+SLACK	0.0	NONE	0.097167	NONE	" " " " "
49	NO2MXV	83.510435	+SLACK	0.0	NONE	NONE	NONE	" " " " "
50	NO2MNV	59.650310	-SLACK	0.0	NONE	NONE	NONE	" " " " "

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H-COAL CASE 1
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
51	NO2MXS	1.096055	+SLACK	0.0	NONE	NONE	NONE	SLACK OF PROPERTIES TO SPECIFICATION
52	NO2MXD	0.0	+SLACK	0.0	NONE	424.524037	NONE	" " " " "
53	NO6MXV	0.0	+SLACK	0.0	NONE	2.435466	NONE	" " " " "
54	NO6MNV	0.0	-SLACK	0.0	NONE	NONE	NONE	" " " " "
55	NO6MXS	0.0	+SLACK	0.0	NONE	NONE	NONE	" " " " "
56	NO6MXD	0.0	+SLACK	0.0	NONE	520.274513	NONE	" " " " "
57	RFOMXV	16.430667	+SLACK	0.0	NONE	NONE	NONE	" " " " "
58	RFOMNV	106.799333	-SLACK	0.0	NONE	NONE	NONE	" " " " "
59	RFOMXS	1.349779	+SLACK	0.0	NONE	NONE	NONE	" " " " "
60	RFOMXD	0.322863	+SLACK	0.0	NONE	NONE	NONE	" " " " "
61	TF1MXV	47.081111	+SLACK	0.0	NONE	NONE	NONE	" " " " "
62	TF1MNV	32.918889	-SLACK	0.0	NONE	NONE	NONE	" " " " "
63	TF1MXS	1.078630	+SLACK	0.0	NONE	NONE	NONE	" " " " "
64	TF1MXN	0.382084	+SLACK	0.0	NONE	NONE	NONE	" " " " "
65	TF1MXD	0.147086	+SLACK	0.0	NONE	NONE	NONE	" " " " "
66	TF1X65	0.0	+SLACK	0.0	NONE	21.396011	NONE	" " " " "
67	TF1X95	0.0	+SLACK	0.0	NONE	NONE	NONE	" " " " "
68	B1DYLD	0.0	-SLACK	0.0	NONE	30.610448	NONE	
69	CH1YLD	0.0	-SLACK	0.0	NONE	31.142313	NONE	
70	D1DYLD	0.0	-SLACK	0.0	NONE	33.188667	NONE	
71	D1FYLD	0.0	-SLACK	0.0	NONE	33.188667	NONE	
72	G1DYLD	0.0	-SLACK	0.0	NONE	22.490662	NONE	
73	G1FYLD	0.0	-SLACK	0.0	NONE	25.417528	NONE	
74	G1HYLD	0.0	-SLACK	0.0	NONE	33.188667	NONE	
75	HHDYLD	0.0	-SLACK	0.0	NONE	40.561655	NONE	
76	H1DYLD	0.0	-SLACK	0.0	NONE	34.566964	NONE	
77	H1HYLD	0.0	-SLACK	0.0	NONE	41.921451	NONE	
78	L1DYLD	0.0	-SLACK	0.0	NONE	40.164249	NONE	
79	L1HYLD	0.0	-SLACK	0.0	NONE	40.770998	NONE	
80	N1FYLD	0.0	-SLACK	0.0	NONE	41.117491	NONE	
81	R11YLD	0.0	-SLACK	0.0	NONE	42.334596	NONE	
82	R16YLD	0.0	-SLACK	0.0	NONE	42.111038	NONE	

H-COAL CASE 1
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
****	STRUCTURAL VARIABLES							
1	TOTPDV	1429.725234	IN BDS	0.0	NONE	NONE	1.000000	MM$ TOTAL PRODUCT VALUE
2	TOTFDC	1600.000000	IN BDS	0.0	NONE	NONE	-1.000000	MM$ TOTAL FEED COST
3	TOTOPC	5.023331	IN BDS	0.0	NONE	NONE	-1.000000	M$ OPERATING COST (CHEMICALS, WATER)
4	TOTIIC	55.602257	IN BDS	0.0	NONE	NONE	-2.160000	
5	TOTIC2	88.358579	IN BDS	0.0	NONE	NONE	-2.160000	
6	TOTRCC	4.672995	IN BDS	0.0	NONE	NONE	-2.160000	MM$ ROYALT. AND CAT. COST OF EXT. REFIN.
7	TOTRC2	4.008546	IN BDS	0.0	NONE	NONE	-2.160000	MM$ ROYALT. AND CAT. COST OF NEW REFIN.
8	TOTCAP	60.274852	IN BDS	0.0	NONE	NONE	0.0	
9	ADDCAP	92.367125	IN BDS	0.0	NONE	NONE	0.0	
10	TOTLPG	2.311783	IN BDS	0.0	NONE	NONE	0.0	MBBL TOTAL LPG PRODUCED
11	TOTNLG	19.792874	IN BDS	0.0	NONE	NONE	0.0	MBBL TOTAL NON LEAD GASOLINE PRODUCED
12	MOLNLG	53.222201	IN BDS	0.0	NONE	NONE	0.0	
13	TOTNO2	17.895093	IN BDS	0.0	NONE	NONE	0.0	MBBL TOTAL NO2 FUEL OIL PRODUCED
14	MLBNO2	5.483057	IN BDS	0.0	NONE	NONE	0.0	
15	TOTNO6	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL TOTAL NO6 FUEL OIL PRODUCED
16	MLBNO6	0.0	IN BDS	0.0	NONE	NONE	0.0	
17	TOTSUL	0.010213	IN BDS	0.0	NONE	NONE	0.0	MLT TOTAL SULFUR PRODUCED
18	TOTNH3	0.033971	IN BDS	0.0	NONE	NONE	0.0	MT " AMMONIA PRODUCED
19	TOTCOK	0.000000	IN BDS	0.0	NONE	NONE	0.0	MT " COKE FROM COKER PRODUCED
20	TOTRFO	4.107667	IN BDS	0.0	NONE	NONE	0.0	MBBL " REFINERY FUEL OIL
21	MLBRFO	1.350601	IN BDS	0.0	NONE	NONE	0.0	
22	HTVLPG	9.574083	IN BDS	0.0	NONE	NONE	0.0	MMMBTU HEATING VALUE OF PRODUCTS
23	HTVNLG	104.045831	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
24	HTVNO2	100.693701	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
25	HTVNO6	0.0	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
26	HTVTF1	28.461190	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
27	HTVSUL	0.091276	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
28	HTVNH3	0.656857	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
29	HTVCOK	0.000000	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " "
30	TOTHTV	243.522938	IN BDS	0.0	NONE	NONE	0.0	MMMBTU TOTAL HEATING VALUE OF PRODUCTS
31	HTVFED	287.170000	IN BDS	0.0	NONE	NONE	0.0	MMMBTU " " " " FEED
32	HTVRFO	23.988773	IN BDS	0.0	NONE	NONE	0.0	MMMBTU HEATING VALUE OF REF. FUEL OIL
33	LHVFUL	23.971669	IN BDS	0.0	NONE	NONE	0.0	MMMBTU LOW HEAT VALUE OF FUEL CONSUMP.
34		0.0	AT MIN	0.0	NONE	0.0	0.0	
35	TOTTF1	5.000000	AT MIN	5.000000	NONE	33.188666	0.0	MBBL TOTAL TURBINE FUEL PRODUCED
36	MLBTF1	1.541914	IN BDS	0.0	NONE	NONE	0.0	
37	NHTTOT	14.504000	IN BDS	0.0	NONE	NONE	-0.000001	MBBL NAPHTHA HYDROTR. CAPACITY (EXIST.)
38	ADDNHT	0.0	FIXED	0.0	0.0	0.374545	-0.000002	MBBL ADDITIONAL NAPHTHA H-TR CAPACITY
39	REFTOT	11.796113	IN BDS	0.0	NONE	NONE	-0.000001	MBBL REFORMER CAPACITY (EXIST.)
40	ADDREF	0.0	FIXED	0.0	0.0	1.904689	-0.000002	MBBL ADDITIONAL REFORMER CAPACITY
41	DHTTOT	0.0	AT MIN	0.0	NONE	0.000001	-0.000001	MBBL DISTILLATE H-TR. CAPACITY (EXIST.)
42	ADDDHT	0.0	FIXED	0.0	0.0	1.010666	-0.000002	MBBL ADDIT. DIST. H-TREATER CAPACITY
43	FCCTOT	0.0	AT MIN	0.0	NONE	0.000001	-0.000001	MBBL FLUID CAT CRACKER CAPACITY (EXIST.)
44	ADDFCC	0.0	FIXED	0.0	0.0	2.096930	-0.000002	MBBL ADDIT. FCC CAPACITY
45	REKTOT	0.0	AT MIN	0.0	NONE	0.000001	-0.000001	MBBL RESID COKER CAPACITY (EXIST.)
46	ADDREK	0.0	FIXED	0.0	0.0	2.473202	-0.000002	MBBL ADDIT. COKER CAPACITY
47	GHCTOT	3.972351	IN BDS	0.0	NONE	NONE	-0.000001	MBBL GAS OIL H-CRACKER CAPACITY (EXIST.)
48	ADDGHC	0.0	FIXED	0.0	0.0	5.750785	-0.000002	MBBL ADDIT. H-CRACKER CAPACITY
49	ALKTOT	0.0	AT MIN	0.0	NONE	0.000001	-0.000001	MBBL ALKYLATION CAPACITY (EXIST.)
50	ADDALK	0.0	FIXED	0.0	0.0	2.452466	-0.000002	MBBL ADDIT. ALKYLATION CAPACITY
51	H2PPLT	37.856016	IN BDS	0.0	NONE	NONE	0.0	MMSCF HYDROGEN PLANT CAPACITY (EXIST.)

Exhibit 5-A (Cont'd)

H-COAL CASE 1
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(OJ)	INPUT COST(CJ)	
52	ADDH2P	0.0	FIXED	0.0	0.0	0.000002	-0.000002	MMSCF ADDIT. HYDROGEN PLANT CAPACITY
53	POXPLT	0.0	AT MIN	0.0	NONE	2.857993	0.0	MMSCF PARTIAL OXIDATION PLANT CAPACITY
54	COKPUR	0.0	AT MIN	0.0	NONE	5.000000	0.0	
55	H2PPUR	0.0	AT MIN	0.0	NONE	6.854460	0.0	
56	SULPLT	0.010213	IN BDS	0.0	NONE	NONE	0.0	MLT SULFUR PLANT CAPACITY. (EXIST.)
57	ADDSUL	0.0	FIXED	0.0	0.0	0.000002	-0.000002	MLT ADDIT. SULFUR PLANT
58	SWSPLT	222[illegible]382	IN BDS	0.0	NONE	NONE	0.0	MLB SOUR WATER STRIPPER CAPACITY (EXIST.)
59	ADDSWS	0.0	FIXED	0.0	0.0	0.000002	-0.000002	MLB ADDIT. SOUR WATER STRIPPER CAPACITY
60	CONTSW	0.0	AT MIN	0.0	NONE	0.000048	0.0	
61	NH3PLT	0.033971	IN BDS	0.0	NONE	NONE	0.0	MT AMMONIA PLANT CAPACITY (EXIST.)
62	ADDN[illegible]	0.0	FIXED	0.0	0.0	0.000002	-0.000002	MT ADDIT. AMMONIA PLANT CAPACITY
63	CWCPLT	54.789305	IN BDS	0.0	NONE	NONE	0.0	MMGAL COOLING WATER (CIRC) CAPAC. (EXIST.)
64	ADDCWC	0.0	FIXED	0.0	0.0	0.000002	-0.000002	MMGAL ADDIT. COOL. WATER CAPACITY
65	CONPLT	12945.802336	IN BDS	0.0	NONE	NONE	0.0	
66	KWHPLT	377.608995	IN BDS	0.0	NONE	NONE	0.0	MKWH POWER PLANT POWER PRODUCTION
67	S12PLT	3482.232753	IN BDS	0.0	NONE	NONE	0.0	MLB POWER PLT 1250 PSI STEAM PROD. (EXIST.)
68	ADDS12	0.0	FIXED	0.0	0.0	0.000000	-0.000000	MLB ADDIT. POWER PLANT STEAM PRODUCTION
69	KWHPT1	62.981240	IN BDS	0.0	NONE	NONE	0.0	
70	KWHPT2	59.977845	IN BDS	0.0	NONE	NONE	0.0	
71	KWHPT3	97.278708	IN BDS	0.0	NONE	NONE	0.0	
72	KWHPT4	28.863859	IN BDS	0.0	NONE	NONE	0.0	
73	KWHPT5	70.178496	IN BDS	0.0	NONE	NONE	0.0	
74	KWHPT6	58.328847	IN BDS	0.0	NONE	NONE	0.0	
75	TOTCOF	0.0	IN BDS	0.0	NONE	NONE	0.0	
76	TOTCWD	0.0	IN BDS	0.0	NONE	NONE	0.0	
77	TOTBHP	71.226096	IN BDS	0.0	NONE	NONE	0.0	
78	BHPMOT	71.226096	IN BDS	0.0	NONE	NONE	0.0	
79	BHPTUR	0.0	IN BDS	0.0	NONE	NONE	0.0	
80	TOTBTU	11.999834	IN BDS	0.0	NONE	NONE	0.0	
81	C3-ALK	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL C3- UNSAT TO ALKYLATION
82	C4-ALK	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL C4- UNSAT TO ALKYLATION
83	C1PHPT	0.0	FIXED	0.0	0.0	-4.779569	0.0	MMSCF C1 TO HYDROGEN PLANT
84	C2PHPT	0.0	FIXED	0.0	0.0	-8.085733	0.0	MMSCF C2 " " "
85	C3PHPT	0.0	FIXED	0.0	0.0	-8.510690	0.0	MBBL C3 " " "
86	H2PFUL	0.086029	IN BDS	0.0	NONE	NONE	0.0	MMSCF LIGHT GASES TO FUEL
87	C1PFUL	0.396560	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
88	C2PFUL	0.225744	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
89	C3PFUL	0.022913	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
90	IC4FUL	0.005720	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
91	NC4FUL	0.081987	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " "
92	H2UFUL	0.0	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
93	C1UFUL	0.0	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
94	C2-FUL	0.0	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
95	C2UFUL	0.0	IN BDS	0.0	NONE	NONE	0.0	MMSCF " " " "
96	C3-FUL	0.0	AT MIN	0.0	NONE	14.236242	0.0	MBBL " " " "
97	C3UFUL	0.0	AT MIN	0.0	NONE	10.634435	0.0	MBBL " " " "
98	C4-FUL	0.0	AT MIN	0.0	NONE	16.177031	0.0	MBBL " " " "
99	TOTC3-	0.0	IN BDS	0.0	NONE	NONE	0.0	
100	TOTC3P	0.152754	IN BDS	0.0	NONE	NONE	0.0	
101	TOTC3U	0.0	IN BDS	0.0	NONE	NONE	0.0	
102	TOTIC4	0.063558	IN BDS	0.0	NONE	NONE	0.0	
103	TOTI4U	0.0	IN BDS	0.0	NONE	NONE	0.0	
104	TOTC4-	0.0	IN BDS	0.0	NONE	NONE	0.0	

Exhibit 5-A (Cont'd)

H-COAL CASE 1
TURBINE FUEL NO.1, MAX N=0.25 WT%

	NAME	VALUE	STATUS	MIN VALUE	MAX VALUE	COST OF BND(DJ)	INPUT COST(CJ)	
105	TOTNC4	4.049342	IN BDS	0.0	NONE	NONE	0.0	
106	TOTN4U	0.0	IN BDS	0.0	NONE	NONE	0.0	
107	TOTSFG	0.860294	IN BDS	0.0	NONE	NONE	0.0	
108	TOTH2H	57.469496	IN BDS	0.0	NONE	NONE	0.0	
109	CO1HTR	50.000000	FIXED	50.000000	50.000000	6.781229	0.0	MBBL RAW H-COAL OIL TO HYDROTREATER
110	CH1DIS	51.800000	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. H-COAL OIL TO DISTILLATION
111	H1DHTR	14.504000	IN BDS	0.0	NONE	NONE	0.0	MBBL HVY NAPHTHA FROM H-COAL-DIST TO H-TR.
112	B1DHCR	3.972351	IN BDS	0.0	NONE	NONE	0.0	MBBL H-COAL DIST. BOTTOMS TO H-CRACKER
113	B1DDIS	25.294649	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " SPLITTER
114	G1DFCC	0.0	FIXED	0.0	0.0	-3.368233	0.0	MBBL HVY GAS OIL FROM SPLITTER TO FCC
115	NAHREF	0.0	AT MIN	0.0	NONE	1.359796	0.0	MBBL NAPHTHA TO REFORMER
116	HHDR96	11.796113	IN BDS	0.0	NONE	NONE	0.0	MBBL H-TR. NAPHTHA TO 96 RON REFORMER
117	HHDR10	0.0	AT MIN	0.0	NONE	0.565789	0.0	MBBL " " " 100 " "
118	L1DHPT	1.941334	IN BDS	0.0	NONE	NONE	0.0	MBBL LT. NAPHTHA TO HYDROGEN PLANT
119	L1HHPT	0.0	AT MIN	0.0	NONE	0.606749	0.0	MBBL " " " " "
120	C3-LPG	0.0	AT MIN	0.0	NONE	3.821120	0.0	MBBL C3 & C4 TO LPG
121	C3PLPG	0.129841	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
122	C3ULPG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
123	C4-LPG	0.0	AT MIN	0.0	NONE	8.032460	0.0	MBBL " " " " "
124	IC4LPG	0.057837	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
125	NC4LPG	2.124104	IN BDS	0.0	NONE	NONE	0.0	MBBL " " " " "
126	IC4NLG	0.0	AT MIN	0.0	NONE	2.399940	0.0	MBBL C4 TO GASOLINE
127	NC4NLG	1.893251	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
128	C4-NLG	0.0	AT MIN	0.0	NONE	11.116988	0.0	MBBL " " "
129	AL3NLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL ALKYLATE TO GASOLINE
130	AL4NLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " " "
131	L1DNLG	2.047266	IN BDS	0.0	NONE	NONE	0.0	MBBL H-COAL LT-NAPHTHA TO GASOLINE
132	L1HNLG	0.671327	IN BDS	0.0	NONE	NONE	0.0	MBBL " HYDROCR. LT NAPHT. TO GASOL.
133	N1FNLG	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " FCC NAPHTHA TO GASOLINE
134	H1HNLG	1.620719	IN BDS	0.0	NONE	NONE	0.0	MBBL " H-CRACK. HVY NAPHT. TO GASOL.
135	HHDNLG	2.707687	IN BDS	0.0	NONE	NONE	0.0	MBBL " H-TREAT. " " " "
136	R16NLG	10.852424	IN BDS	0.0	NONE	NONE	0.0	MBBL " 96 RON REFORMATE " "
137	R11NLG	0.000000	IN BDS	0.0	NONE	NONE	0.0	MBBL " 100 " " " "
138	D1DNO2	15.906749	IN BDS	0.0	NONE	NONE	0.0	MBBL " DISTILLATE TO NO2 FUEL OIL
139	G1DNO2	1.988364	IN BDS	0.0	NONE	NONE	0.0	MBBL " GAS OIL TO NO2 FUEL OIL
140	G1HNO2	0.0	AT MIN	0.0	NONE	5.943337	0.0	MBBL " H-CRACK. GAS OIL TO NO2 FO
141	D1FNO2	0.0	AT MIN	0.0	NONE	20.929035	0.0	MBBL " FCC LC GAS OIL TO NO2 FO
142	D1DNO6	0.0	AT MIN	0.0	NONE	4.893485	0.0	MBBL " DISTILLATE TO NO6 FUEL OIL
143	G1DNO6	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " GAS OIL TO NO6 FUEL OIL
144	G1HNO6	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " H-CRACK GAS OIL TO NO6 FO
145	D1FNO6	0.0	AT MIN	0.0	NONE	29.325286	0.0	MBBL " FCC LC GAS OIL TO NO6 FO
146	G1FNO6	0.0	AT MIN	0	NONE	15.137685	0.0	MBBL " " HVY " " " " "
147	D1DRFO	0.0	AT MIN	0.0	NONE	11.622280	0.0	MBBL " DISTILLATE TO REFINERY FUEL
148	G1DRFO	4.107667	IN BDS	0.	NONE	NONE	0.0	MBBL " GAS OIL TO REF. FUE
149	G1HRFO	0.0	AT MIN	0.0	NONE	10.582472	0.0	MBBL " H-CRACK. GAS OIL TO REF. FUEL
150	D1FRFO	0.0	AT MIN	0.0	NONE	9.196061	0.0	MBBL " FCC LC GAS OIL TO REF. FUEL
151	G1FRFO	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " " HVY " " " " "
152	D1DTF1	3.291889	IN BDS	0.0	NONE	NONE	0.0	MBBL " DISTILLATE TO TURB. F FUEL
153	G1DTF1	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " GAS OIL " " "
154	G1HTF1	1.708111	IN BDS	0.0	NONE	NONE	0.0	MBBL " H-CR. GAS OIL " " "
155	D1FTF1	0.0	IN BDS	0.0	NONE	NONE	0.0	MBBL " FCC LC GAS OIL " " "
156	G1FTF1	0.0	AT MIN	0.0	NONE	13.624873	0.0	MBBL " " HVY " " " " "

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SECTION 6

DATA EVALUATION

This section presents the evaluations of the data developed and presented in the previous task sections.

6.1 LITERATURE SURVEY

The report covering the Task I - Literature Survey activity was written and submitted in April, 1980. The earlier Literature Survey report has become an Appendix to this report. A major portion of the data required to complete the subsequent tasks, reported in Sections 3, 4 and 5, was obtained from the Appendix document.

6.2 ON-SITE FUEL PRETREATMENT

The information presented on this subject in Section 3 of this report indicates that problems are anticipated in the water washing pretreatment operation for a fair percentage of coal, oil shale-derived liquids and resids using the conventional electrostatic precipitator equipment and systems. Alternate equipment which can probably be used is available.

Details given in Section 3 indicate the costs of the alternate centrifugal contactor system, both capital and operating, are no greater than for the conventional equipment system. The comparison, summarized from data in Section 3, is as follows:

C-2

Alkali Metals (ppm)	FCI Conventional System ($ Thousand)	Alternate System ($ Thousand)	Difference ($ Thousand)	Percent Difference
To 20	1,680	1,200	480	28
20 to 200	2,050	1,200	850	41
200 to 2000	2,560	1,725	835	32

Alkali Metals (ppm)	Annual Operating Costs Conventional System ($ Thousand)	Alternate System ($ Thousand)	Difference Annual ($ Thousand)	Percent Difference
To 20	588	432	104	26
20 to 200	689	432	205	37
200 to 2000	819	563	204	31

6.3 EXISTING AND NEW REFINERIES TO UPGRADE FUELS

In the following sections, 6.4 through 6.6, the process paths resulting from the linear programming calculation are evaluated for the capital cost of additional process units and the price of turbine fuel. The price of turbine fuel is determined as a means of supporting the profitability of the refinery expansion to meet a 15 percent discounted cash flow. Also, the thermal efficiency of the combined processes and the utilities requirements are determined.

For simplicity of presentation, evaluations for each feed material will be made for both existing and new refinery operations cases.

6.4 SHALE OIL UPGRADING

6.4.1 EXISTING REFINERY TO UPGRADE SHALE OIL

The refinery model combines the petroleum refinery operation with the shale oil refinery operation by blending product streams to meet a given product slate. Additional process units are included where petroleum and shale oil require separate treatment at different severity levels to meet product specifications. Based on a fixed feed of 50,000 BPD of shale oil, and production limits of gasoline, No. 2 fuel oil, No. 6 fuel oil, and turbine fuel, the program finds the most economical process route by reducing petroleum crude feed. The calculations predict required product selling prices at different nitrogen levels and endpoint specifications for turbine fuels in order to establish the impact of turbine fuel quality on the process economics.

A. Refinery LP Output Configuration

The refinery configurations represent economical processes for the combination of petroleum and shale oil when different hydrotreating alternatives are applied. Figure 6-1, Case 1, shows severe hydrotreating of whole shale oil before distillation with further upgrading of the distillation cuts. Figure 6-2, Case 2, shows only mild hydrotreating of whole shale oil with severe hydrotreating of the distillation cuts.

The addition of process units to the base case refinery and the resulting investment costs are described in Section 4.2.2A and 4.2.2D.

The major difference between the two configurations is in the amount of petroleum feed reduction, and the lower nitrogen level of 0.02 wt% in the turbine fuel in Figure 6-1, Case 1. In the combined refinery, Cases 1 and 2, the linear program refinery model was allowed to blend two different turbine fuel types: a distillate turbine fuel product with a 650°F endpoint, and a heavier turbine fuel product with a greater than 1000°F endpoint. To show the influence of the nitrogen limit on the turbine fuel, Figure 6-2,

Case 2, was blended to meet two turbine fuel product specifications, one with a 0.25 wt% limit, and the other with a 1.0 wt% limit of nitrogen. A description of these turbine fuel specifications is shown in Table 6-1.

B. Calculation of Turbine Fuel Prices

Determination of the price of turbine fuel produced from a combined refinery consisting of petroleum and shale oil feed requires definition of several basic operating conditions. Little is known about the demand factors that would affect a future turbine fuel market. Therefore, the following operating conditions were set up for an expanded refinery with shale oil upgrading to arrive at an acceptable relative price for turbine fuel:

(1) The amount of gasoline and No. 2 fuel oil is held constant since the market for these fuels does not change, disregarding normal seasonal variations.

(2) 8,000 BPD of No. 6 fuel oil are produced.

(3) 20,000 BPD of turbine fuel is produced in the combined refinery cases.

(4) All product prices, except turbine fuel, stay the same. Turbine fuel required selling price supports the profitability of the refinery expansion to meet a 15 percent discounted cash flow rate of return.

(5) The feed of shale oil is fixed at 50,000 BPD, while crude oil is reduced to meet the given product slate.

The results of the different refinery configurations producing several grades of turbine fuel are shown in Table 6-4. Table 6-2 represents capital cost data for the combined refinery with severe hydrotreating before distillation, Case 1. Table 6-4 includes the calculated required selling prices for turbine fuels TF1 and TF3 for Case 1. TF1 and TF3 are described in Table 6-1.

Table 6-3 represents capital cost data for the combined refinery with mild hydrotreating before distillation, Case 2. Table 6-4 also contains the calculated required selling prices for turbine fuels TF1, T11, TF3, and T13 for Case 2. T11 and T13 are described in Table 6-1.

In order to formulate Table 6-4, daily feed, operating costs and product values are obtained as computer outputs for each optimum mode of operation. The total daily required products selling price or revenue is manually calculated by adding feed and operating costs to a "capital recovery factor." This factor, amounting to 35% of the FCI, is based on the following:

(1) 15% DCF rate of return
(2) 50% income tax
(3) 10% investment tax credit
(4) double declining balance depreciation with 16 years useful life
(5) 20-year operation
(6) 4% of FCI as annual maintenance costs
(7) 2.5% of FCI as annual property taxes and insurance costs
(8) Allowance for spare parts inventory
(9) 330 operating days per year

The product values, exclusive of those for turbine fuel, are deducted from the required revenue. The difference is the revenue which is a portion of turbine fuel sales. The sum of this revenue difference and the operating margin for the existing basic petroleum refinery as calculated in Section 4, Table 4-3, is the amount for which the turbine fuel must be sold

so that the basic refinery operation profit is not penalized by the added capital intensive shale oil facilities and feed.

The product slates for Cases 1 and 2 are essentially unchanged for the several turbine fuel quality specifications and are shown in Tables 6-5 and 6-6.

C. Evaluation of Turbine Fuel Prices Versus Turbine Fuel Quality

The turbine fuels, described in subparagraph 6.4.1A and Table 6-1, were chosen to represent different grades of turbine fuel quality with respect to distillation and endpoint, viscosity and nitrogen content. By producing these different grades in the combined refinery, turbine fuel quality versus price can be evaluated. Because of the many different blending stocks that may be chosen and varied for each class of turbine fuel, only the overall cost calculations including capital cost, feed cost, and product value can give a relative turbine fuel price versus quality change.

An evaluation of the turbine fuel required selling price calculations in Table 6-4 indicates the factors that affect price are as follows:

(1) The fixed capital investment for additional process units has a major effect on turbine fuel price.

(2) The difference between feed cost and product value for the different grades of turbine fuel has a less significant effect on turbine fuel price.

In Case 1, severe hydrotreating of whole shale oil reduces the nitrogen below fuel specification. Thus, a change in nitrogen limit has no effect. A change to a heavier endpoint fuel improves the overall economics of the refinery resulting in a turbine fuel price reduction of about 2%.

In Case 2, mild hydrotreating of whole shale oil, a change to heavier endpoint fuel has a major effect in decreasing capital cost of new process units. This results in a turbine fuel price reduction of about 6%.

Also in Case 2, the nitrogen level was varied for light and heavy turbine fuels which showed a major reduction in capital cost at the higher nitrogen level. The result is a turbine fuel required selling price reduction of about 13% for distillate fuel (TF1) and about 10% for heavier fuel (TF3).

These lower turbine fuel prices in Case 2 result from deletion of the mid-distillate hydrotreater for the higher nitrogen content turbine fuel. This reduces the hydrogen demand and improves the overall economics. When turbine fuel specification is changed from light to heavy fuel, the expansion of the hydrocracker is not required which reduces new refinery cost.

An evaluation of turbine fuel prices versus turbine fuel quality is shown in Figures 6-3 and 6-4. These curves are a plot of the calculated turbine fuel price versus the nitrogen level contained in the turbine fuel. The curves show two levels of maximum nitrogen content:

(1) Fuels having 0.25 wt% nitrogen as the maximum acceptable nitrogen content for present day gas turbines, and

(2) Fuels having 1.0 wt% nitrogen as the maximum acceptable nitrogen content for gas turbines with combustion modification or possibly flue gas treatment.

These curves represent the range of nitrogen content in turbine fuels for present and future gas turbine combustions. An evaluation of turbine fuel price versus endpoint specification for turbine fuels is shown in Figure 6-4 for Cases 1 and 2. These curves are a plot of the calculated turbine fuel price versus a distillate type and a wide boiling range turbine fuel. The properties of the distillate and wide-range turbine

fuels are shown in the table on Figure 6-4. To produce the wide-range turbine fuel, about 11 percent heavy resid is blended with the fuel which results in a lower gravity and slightly higher viscosity of product. In these calculated cases, the sulfur specification of 0.7 wt% limited the fraction of heavy resid that was blended into turbine fuel.

The foregoing description and discussion of the LP program application, basis and methods of calculation under these subsections 6.4.1A, B and C pertaining to the existing petroleum refinery with normal petroleum crude plus shale oil feed, also applies to the subsequent subsections covering the alternate synfuel feeds. Accordingly, the repetition of applicable similar descriptions and discussions will be avoided.

D. Thermal Efficiency

The thermal efficiencies of the shale oil plus existing petroleum refining for Cases 1 and 2 and each of the turbine fuel specifications are shown in Table 6-7. The thermal efficiency for Case 1 turbine fuels TF1 and TF3 is 89.1% for both fuels. The thermal efficiencies for Case 2 turbine fuels TF1, T11 and TF3, T13 range from 88.3% to 89.4%.

E. Utilities

The utilities requirements shown in Table 6-8 are based on providing 1,250 psig steam for driving letdown turbines to provide power requirements and low level process steam. Fuel is provided from refinery fuel gas and fuel oil generated internally for firing heaters and boiler facilities. Cooling water, condensate, and sour water stripping facilities are also provided.

6.4.2 NEW SHALE OIL REFINERY

Unlike the process calculation for the combined shale oil plus petroleum refinery, the stand-alone shale oil refinery is not given a product slate to meet, with the exception of 5,000 BPD of turbine fuel which has to be produced. LPG, gasoline, No. 2 and No. 6 Fuel Oil will be produced and blended to maximize the product value. With a fixed feed of 50,000 BPD shale oil, the program finds the most economical process route, based on process yields and severity levels for the treatment of the different distillation fractions.

The refinery has to provide its own fuel for utility production. Hydrogen is produced from light gases from the refinery, but a unit for the partial oxidization of coke to hydrogen is included to provide the hydrogen shortfall which cannot be produced from refinery streams.

To determine the impact of turbine fuel quality on the process economics, the linear program model was allowed to blend to different turbine fuel specifications, as described in the combined refinery cases.

A. Resulting Refinery Linear Programming Configuration

The refinery configurations represent economical process routes for upgrading shale oil in a new refinery when different hydrotreating methods are applied. The difference between the two configurations, Figures 6-5 and 6-6, is the degree of hydrotreating before and after distillation. In Figure 6-5, Case 1, severe hydrotreating at high pressure and low space velocity is employed to hydrodenitrify the whole shale oil feed to a nitrogen level of about 500 ppm (wt). The result is pgrading of whole shale oil from an API of 21.4 to 38.0 degrees, a liquid resembling petroleum crude. The resulting 650°+F fraction is an excellent feed for FCC process or hydrocracking processes.

In Figure 6-6, Case 2, hydrotreating after distillation of individual fractions takes place at high pressure and low space velocity to reduce the nitrogen to the level required to prevent poisoning and deactivation of the catalyst in subsequent processing units. Hydrotreating of the 650°F+ fraction for FCC feed was less severe than for hydrocracking.

No. 6 fuel oil was not produced in the calculated Cases 1 and 2, due to the small amount of high boiling fraction available and the demand for refinery fuel oil for utility production. In both of the calculated Cases 1 and 2, gasoline production was maximized to increase total product value.

Heavy turbine fuel (TF3) was not produced in Figure 6-5, Case 1, because of the small amount of higher boiling point fractions available for blending. Also, no high nitrogen (1 wt%) turbine fuel (T11) was produced because of the severe hydrotreating of whole shale oil which reduced the nitrogen level below 0.25 wt%. Thus, only a distillate turbine fuel (TF1) with a 650°F endpoint and less than 0.25 wt% nitrogen was produced.

In Figure 6-6, Case 2, the linear programming was allowed to blend two different turbine fuel types: a distillate turbine fuel with a 650°F endpoint, and a heavier turbine fuel. Also, both turbine fuels contained up to 1 wt% nitrogen. Thus, the turbine fuel cases were the same as produced in the combined petroleum/shale oil facility, namely, TF1, T11, TF3, T13.

B. Calculation of Turbine Fuel Prices

To determine a value for turbine fuel for the shale oil refinery, the complete calculation was based on forcing the turbine fuel production of 5,000 BPD at zero value. After deducting the daily capital recovery, operating cost and feed cost from the product value (excluding turbine fuel), a revenue margin was left which had to be supported by the

turbine fuel price. This price represents the value of turbine fuel to a shale oil refinery forced to produce 5,000 BPD of turbine fuel. The turbine fuel price is based on selling all other products with petroleum specifications at the market prices prevailing for comparable petroleum products.

The foregoing description and discussion, relative to the LP program application and the basis and calculation methods pertaining to the new stand-alone shale oil refinery contained in subsections 6.4.2A and B, also apply to the subsequent subsections covering the remainder of the new refineries for upgrading the individual synfuels.

Table 6-9 presents capital cost data for the new shale oil refinery with severe hydrotreating before distillation, Case 1. Table 6-10 presents the calculated turbine fuel prices for turbine fuels TF1, T11, TF3, and T13 for Cases 1 and 2. Table 6-11 presents capacity and capital cost data for the new shale oil refinery with mild hydrotreating before distillation, Case 2.

C. Evaluation of Turbine Fuel Prices Versus Turbine Fuel Quality

The evaluation of the turbine fuel price calculations, as shown in table 6-10, indicates the key factors that affect required selling price are as follows:

(1) The data in Table 6-10, Case 1, TF1, severe hydrotreating before distillation, indicates a high turbine fuel price is required to provide the 15% discounted cash flow profit level of the new shale oil refinery. This price is about $116, or about 3.5 times the turbine fuel required selling price from the combined shale oil plus petroleum refinery.

The factor that most affects this price difference is the high capital investment cost for the new shale oil refinery. Unlike the existing refinery, where petroleum crude feed is reduced to allow existing process units to be used for shale oil refining, all units must be built new. Also, a partial oxidation unit adds to the total cost for hydrogen production.

(2) The data in Table 6-10, Case 2, TF1, T11, TF3, T13, severe hydrotreating after distillation, indicates a slightly lower capital investment cost than for Case 1 which results in a lower turbine fuel required selling price range of \$98 to \$103. The major cause for the turbine fuel price range is the variation in capital investment for the FCC, partial oxidation and power plant units.

An evaluation of turbine fuel prices versus turbine fuel quality is shown in Figures 6-7 and 6-8. These curves are a plot of the calculated turbine fuel price versus the nitrogen level contained in the turbine fuel. The curves show two levels of maximum nitrogen content:

a. 0.25 wt% nitrogen as the maximum acceptable nitrogen content for present day gas turbine fuels, and

b. 1.0 wt% nitrogen as the maximum acceptable nitrogen content for gas turbines with combustion modification or possible flue gas treatment.

These curves represent the range of nitrogen content in turbine fuels for present and future gas turbine combustors.

An evaluation of turbine fuel price versus endpoint specification for turbine fuels is shown in Figure 6-8 for Cases 1 and 2.

These curves are a plot of the calculated turbine fuel price versus a distillate type and a wide boiling range type of turbine fuel. The properties of the distillate and wide range turbine fuels are shown in the table below Figure 6-8.

To produce the wide range turbine fuel, heavy fuel having an endpoint over 650°F is blended with distillate fuels, which results in a lower gravity and slightly higher viscosity product. In these calculated cases, the sulfur specification of 0.7 wt% limited the fraction of heavy fuel that was blended into turbine fuel.

D. Thermal Efficiency

The thermal efficiencies of the new shale oil refinery for Cases 1 and 2 and each of the turbine fuel specifications are shown in Table 6-12. The thermal efficiencies for case 1 turbine fuels TF1 and TF3 are 76.4% for both fuels, while the thermal efficiencies for Case 2 turbine fuels TF1, T11, and TF3, T13 range from 72.1 to 73.2%.

E. Utilities

The utilities requirements shown in Table 6-13 are based on providing a 1,250 psig steam plant for driving letdown turbines to provide power requirement and lower pressure process steam. Fuel is provided from refinery fuel gas and fuel oil generated internally for firing heaters and boiler facilities. Cooling water, condensate, and sour water stripping facilities are also provided.

Table 6-1 - Description of Turbine Fuel Specifications

TF1 is a turbine fuel with an endpoint spec of 650°F and a maximum of 0.25 wt% of nitrogen.

T11 has the same property spec as TF1 except the nitrogen limit is raised to 1 weight percent.

TF3 is a turbine fuel with an endpoint above 1000°F and a maximum of 0.25 wt% of nitrogen.

T13 has the same property spec as TF3 except the nitrogen limit is raised to 1 weight percent.

Turbine Fuel 1 (TF1): Distillate Fuel

Distillation EP		650°F
Viscosity	max	5.8 cst at 100°F
Viscosity	min	1.8 cst " "
Gravity	max	337.8 lb/bbl
Sulfur	max	0.7 wt%
Nitrogen	max	0.25 wt%

Turbine Fuel 1 (T11)
like TF1, but relaxed Nitrogen Specification

Nitrogen	max	1.0 wt%

Turbine Fuel 2 (TF2): Distillate Fuel
like TF1, but wider viscosity range allowed

Distillation EP		Below 1000°F
Viscosity	max	30.0 cst at 100°F
Viscosity	min	1.8 cst " "
Gravity	max	337.8 lb/bbl
Sulfur	max	0.7 wt%
Nitrogen	max	0.25 wt%

Turbine Fuel 2 (T12): Distillate Fuel
like TF2, but relaxed Nitrogen Specification

Nitrogen	max	1.0 wt%

Turbine Fuel 3 (TF3): Heavy (Residual) Fuel

Distillation EP		Above 1000°F
Viscosity	max	160 cst at 100°F
Viscosity	min	1.8 cst " "
Gravity	max	337.8 lb/bbl
Sulfur	max	0.7 wt%
Nitrogen	max	0.25 wt%

Turbine Fuel 3 (T13)
like TF3, but relaxed Nitrogen Specification

Nitrogen	max	1.0 wt%

Turbine Fuel 4 (TF4): Heavy (Residual) Fuel
like TF3, but wider viscosity range allowed

Distillation EP		Above 1000°F
Viscosity	max	900 cst at 100°F
Viscosity	min	1.8 cst " "
Gravity	max	337.8 lb/bbl
Sulfur	max	0.7 wt%
Nitrogen	max	0.25 wt%

Table 6-2 - Capacity and Capital Cost Data, Petroleum Plus Shale Oil Refinery, Case 1

	Unit Capacities, BPD			Fixed Capital Investment ($ Million)	
	Existing	Equipment Additions			
Process Units	Refinery	TF1	TF3	TF1	TF3
Crude Unit	200,000	—	—	—	—
Vacuum Distillation	75,000	—	—	—	—
Fluid Catalytic Cracker	50,000	10,916	11,148	19.6	19.9
Hydrocracker	10,300	—	—	—	—
Coker	12,500	—	—	—	—
Naphtha Hydrotreater	61,000	—	—	—	—
Atm Gas Oil Hydrotreater	22,000	—	—	—	—
Reformer	49,000	—	—	—	—
Alkylation	8,000	—	—	—	—
Shale Oil L.P. Hydrotreater	—	50,000	50,000	44.7	44.7
Shale Oil H.P. Hydrotreater	—	49,430	49,430	114.0	114.0
Shale Oil Distillation	—	52,100	52,100	14.0	14.0
Hydrogen Plant, million SCFD	—	83.8	83.2	29.6	29.4
Sulfur Recovery Plant, long ton/day	135	5	—	0.5	—
Ammonia Recovery from Waste Water, ton/day NH_3	17	190	190	7.4	7.4
Sour Water Stripper, M lb/day	5,300	1,126	1,052	0.8	0.8
Cooling Water System, M gal/day	196,000	70,500	71,700	2.6	2.7
Steam/Power Plant, M lb/day, 1250 psig steam	15,100	—	—	—	—
Total Additional FCI				233.2	232.9

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Table 6-3 - Capacity and Capital Cost Data, Petroleum Plus Shale Oil Refinery, Case 2

Process Units	Unit Capacity, BPD					Fixed Capital Investment ($ Million)			
	Existing Refinery	Equipment Additions							
		TF1	T11	TF3	T13	TF1	T11	TF3	T13
Crude Unit	200,000	—	—	—	—	—	—	—	—
Vacuum Distillation	75,000	—	—	—	—	—	—	—	—
Fluid Catalytic Cracker	50,000	16,891	14,293	15,003	14,187	26.6	23.7	24.5	23.5
Hydrocracker	10,300	4,103	—	—	—	25.9	—	—	—
Coker	12,500	—	—	—	—	—	—	—	—
Naphtha Hydrotreater	61,000	—	—	—	—	—	—	—	—
Atm Gas Oil Hydrotreater	21,670	—	—	—	—	—	—	—	—
Reformer	49,000	—	—	—	—	—	—	—	—
Alkylation	8,000	—	—	—	—	—	—	—	—
Shale Oil L.P. Hydrotreater	—	50,000	50,000	50,000	50,000	44.7	44.7	44.7	44.7
Shale Oil Distillation	—	49,430	49,430	49,430	45,960	13.4	13.4	13.4	12.6
Shale Oil Naphtha Hydrotreater	—	1,977	1,977	1,977	1,977	16.7	16.7	16.7	16.7
Shale Oil Distillate Hydrotreater	—	8,807	—	8,950	—	16.7	—	17.4	—
Shale Oil Heavy Gas Oil Hydrotreater	—	25,680	25,900	25,460	22,920	45.3	45.5	45.1	42.3
Hydrogen Plant, million SCFD	—	65.5	50.9	66.3	44.0	24.9	20.9	25.1	18.8
Sulfur Recovery Plant, Long ton/day	135	18	25	16	24	1.3	1.6	1.2	1.6
Ammonia Recovery from Waste Water, ton/day NH_3	17	138	121	137	116	6.1	5.6	6.1	5.5
Sour Water Stripper, M lb/day	5,300	1,719	1,387	1,573	1,321	1.2	1.0	1.1	1.0
Cooling Water System, M gal/day	196,000	63,700	55,600	63,800	48,800	2.4	2.2	2.4	2.0
Steam/Power Plant, M lb/day, 1250 psig steam	15,100	1,022	1,448	1,080	1,479	8.7	11.5	9.1	11.7
Total Additional FCI						233.9	186.8	206.8	180.4

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Table 6-4 - Turbine Fuel Selling Prices, Petroleum Crude Plus Shale Oil Refinery, 20,000 BPD Turbine Fuel Produced, $ per Day

	Case 1		Case 2			
	Two-Stage Hydrotreating		Single-Stage Hydrotreating			
	0.25% Nitrogen Content		0.25% Nitrogen Content		1.0% Nitrogen Content	
Item	650°F Endpoint (TF1)	1000°F+ Endpoint (TF3)	650°F Endpoint (TF1)	1000°F+ Endpoint (TF3)	650°F Endpoint (T11)	1000°F+ Endpoint (T13)
Fixed Capital Investment for Additional Process Units	233,200,000	232,900,000	233,900,000	206,800,000	186,800,000	180,400,000
Add: Offsite Facilities $\frac{\text{FCI Add'tl Proc. Units} \times 0.30}{0.70}$	99,900,000	99,800,000	100,200,000	88,600,000	80,100,000	77,300,000
Royalties and Catalyst	11,642,000	11,650,000	8,594,000	8,373,000	7,527,000	7,129,000
Total Additional Capital Investment	344,742,000	344,350,000	342,694,000	303,773,000	274,427,000	264,829,000
Daily Capital Recovery (Total Add'tl Cap x 0.0010606*)	365,633	365,218	363,451	322,182	291,057	280,876
Add: Feed Cost	6,189,300	6,139,000	6,279,600	6,163,200	6,152,200	6,153,900
Operating Cost	61,019	60,942	60,810	60,268	60,688	59,918
Total Daily Required Revenue	6,615,952	6,565,160	6,703,871	6,545,650	6,503,945	6,494,694
Deduct: Product Values (Exclusive of Turbine Fuel)	6,656,700	6,616,900	6,721,900	6,606,300	6,612,200	6,622,800
Revenue Margin	- 40,748	- 51,740	- 18,028	- 60,650	- 108,254	- 128,106
Add: Operating Margin of Existing Refinery before Addition of Synfuel Upgrading	702,204	702,204	702,204	702,204	702,204	702,204
Turbine Fuel Required Daily Revenue	661,456	650,464	684,175	641,554	593,949	574,098
Minimum Selling Price Per Barrel Turbine Fuel	33.07	32.52	34.21	32.08	29.70	28.70

* Capital Recovery Factor, stream day basis: $\frac{0.35}{330 \text{ operating days per year}} = 0.0010606$

Table 6-5 - Product Slate, Shale Oil Plus Petroleum Refinery, Case 1

Item	Rate	TF1	TF3
Feed			
Petroleum Crude	M BPD	164.64	162.97
Shale Oil	M BPD	50.0	50.0
Products			
LPG	M BPD	14.4	12.9
Gasoline	"	108.1	108.1
No. 2 fuel oil	"	53.8	53.8
No. 6 fuel oil	"	8.0	8.0
Turbine fuel	"	20.0	20.0
Sulfur	M LTPD	.140	.138
Ammonia	M TPD	.207	.207
Coke	M TPD	.653	.507

Table 6-6 - Product Slate, Shale Oil Plus Petroleum Refinery, Case 2

Item	Rate	Turbine Fuels			
		TF1	T11	TF3	T13
Feed					
Petroleum Crude	M BPD	167.65	163.4	163.78	163.46
Shale Oil	M BPD	50.0	50.0	50.0	50.0
Products					
LPG	M BPD	17.3	13.6	12.8	13.5
Gasoline	"	108.1	108.1	108.1	108.1
No. 2 fuel oil	"	53.8	53.3	53.8	53.8
No. 6 fuel oil	"	8.0	8.0	8.0	8.0
Turbine fuel	"	20.0	20.0	20.0	20.0
Sulfur	M LTPD	.153	.160	.151	.159
Ammonia	M LTPD	.155	.138	.154	.133
Coke	M TPD	.695	.668	.570	.643

Table 6-7 - Thermal Efficiencies of Shale Oil Plus Existing Petroleum Refinery

Item	Millions Btu/D Case 1 (TF1)	(TF3)	Case 2 (TF1)	(TI1)	(TF3)	(TI3)
Total Heating Value Feed	1250.4	1240.8	1267.3	1243.3	1245.4	1243.6
Total Heating Value Products	1114.3	1105.2	1128.7	1108.5	1106.3	1111.7
Thermal Efficiency, %	89.1	89.1	89.0	89.2	88.8	89.4

Table 6-8 - Total Utilities Requirement, Shale Oil Plus Existing Petroleum Refinery (Computer Output)

Unit	Usage Rate Case 1	Case 2
Sour water stripping	6388 M lb/D (533 gpm)	6800 M lb/D (567 gpm)
Cooling water circulation	267 MM gal/D (185,415 gpm)	253.9 MM gal/D (176,370 gpm)
Power generation	1538 M kWh/D (64,080 kW)	(1550) M kWh/D (64,560 kW)
Fuel consumption	96 MMM Btu/D	95 MMM Btu/D

Table 6-9 - Fixed Capital Investment, New Shale Oil Refinery, Case 1, Product: Turbine Fuel 1 (TF1)

Capacity, BPD	Process Unit	$ Million
50,000	L.P. Hydrotreater	44.7
49,480	H.P. Hydrotreater	114.0
52,100	Distillation	14.0
17,500	FCC	27.3
5,502	Naphtha Hydrotreater	34.1
5,502	Reformer	10.8
3,034	Alkylation Plant	9.3
50.56 MM SCFD	Hydrogen Plant	20.8
69.261 MM SCFD	Partial Oxid. Plant	121.3
51 LTPD	Sulfur Plant	2.7
192 TPD	Ammonia Plant	7.4
13,558 M lb/D	Sour Water Stripper	6.1
136,000 M gal/D	Cooling Water Plant	4.5
17,136 M lb/D	Power Plant	82.9
		500

Table 6-10 - Turbine Fuel Selling Prices, Shale Oil Refinery, 5,000 BPD Turbine Fuel Produced, $ Per Day

Item	Case 1	Case 2			
	TF1 (0.25% N) 650°F EP	TF1 (0.25% N) 650°F EP	TF3 (0.25% N) 900°F EP	TI1 (1.0% N) 650°F EP	TI3 (1.0% N) 900°F EP
Fixed Capital Investment for Process Units	499,900,000	496,840,000	485,860,000	476,300,000	478,500,000
Add: Offsite Facilities FCI Add'tl Proc. Units x 0.30 / 0.70	214,240,000	212,931,420	208,225,710	204,128,570	205,071,420
Royalties and Catalyst	14,966,000	12,276,000	11,952,000	11,647,000	11,677,000
Total Capital Investment	729,106,000	722,047,420	706,037,710	692,075,570	695,248,420
Daily Capital Amortization (Total Add'tl Cap x 0.0010606)	773,290	765,803	748,824	734,015	737,380
Add: Feed Cost	1,280,130	1,289,175	1,287,450	1,286,400	1,287,200
Operating Cost	32,808	34,751	34,492	33,936	34,119
Total Daily Cost	2,086,228	2,089,729	2,070,766	2,054,351	2,058,699
Deduct: Product Values (Exclusive of Turbine Fuel)	1,506,450	1,574,129	1,581,108	1,548,600	1,569,700
Turbine Fuel Required Daily Revenue	579,778	515,600	489,658	505,751	488,999
Minimum Selling Price Per Barrel Turbine Fuel	115.96	103.12	97.93	101.15	97.80

Table 6-11 - Capacity and Capital Cost Data, New Shale Oil Refinery, Case 2

	Unit Capacity, BPD				Fixed Capital Investment ($ Million)			
	0.25% N		1.0% N					
Process Units	TF1	TF3	T11	T13	TF1	TF3	T11	T13
Low Pressure Hydrotreater	50,000	50,000	50,000	50,000	44.7	44.7	44.7	44.7
Distillation	49,430	49,430	49,430	48,810	13.4	13.4	13.4	13.3
Fluid Catalytic Cracker	31,993	28,758	32,283	30,784	41.6	38.6	41.9	40.5
Naphtha Hydrotreater	1,977	1,977	1,977	1,977	16.7	16.7	16.7	16.7
Distillate Hydrotreater	11,843	12,097	9,279	9,258	20.6	20.9	17.9	17.8
Heavy Gas Oil Hydrotreater	30,182	27,130	30,455	29,008	49.9	46.8	50.2	48.8
Reformer	2,037	2,037	2,037	2,037	5.4	5.4	5.4	5.4
Alkylation	5,068	4,550	5,109	4,800	12.7	11.9	12.7	12.4
Hydrogen Plant, million SCFD	22.5	21.2	22.0	21.3	12.0	11.3	11.9	11.3
Partial Oxidation Plant, million SCFD	90.1	85.5	86.1	83.7	147.7	142.1	142.8	139.8
Sulfur Recovery Plant, long ton/day	63	62	62	62	3.1	3.1	3.1	3.1
Ammonia Recovery from Waste Water, ton/day NH_3	159	152	155	151	6.7	6.5	6.5	6.4
Sour Water Stripper, M lb/day	17,923	17,071	17,187	16,738	8.0	7.4	7.4	7.3
Cooling Water System, M gal/day	145,560	135,900	142,460	137,700	4.7	4.5	4.6	4.5
Steam/Power Plant, M lb/day, 1250 psig steam	24,340	23,066	23,485	22,827	109.7	105.2	104.3	106.7
Total Fixed Capital Investment					496.8	478.5	485.9	476.3

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Table 6-12 - Thermal Efficiencies of New Shale Oil Refinery

	Million Btu/D					
	Case 1		Case 2			
Item	(TF1)	(TF3)	(TF1)	(TF11)	(TF3)	(TI3)
Total Heating Value Feed	335.6	335.6	346.4	344.4	344.1	343.1
Total Heating Value Product	256.5	256.5	249.8	250.6	252.1	251.9
Thermal Efficiency, %	76.4	76.4	72.1	72.8	73.2	73.4

Table 6-13 - Total Utilities Requirement, New Shale Oil Refinery
(Computer Output)

	Usage Rate	
Unit	Case 1	Case 2
Sour water stripping	13,558 M lb/D (1130 gpm)	17,230 M lb/D (1435 gpm)
Cooling water circulation	136 MM gal/D (94,440 gpm)	141 MM gal/D (97,915 gpm)
Power generation	1183 M kWh/D (42,290 kW)	1435 M kWh/D (59,790 kW)
Fuel consumption	45 MMM Btu/D	49 MMM Btu/D

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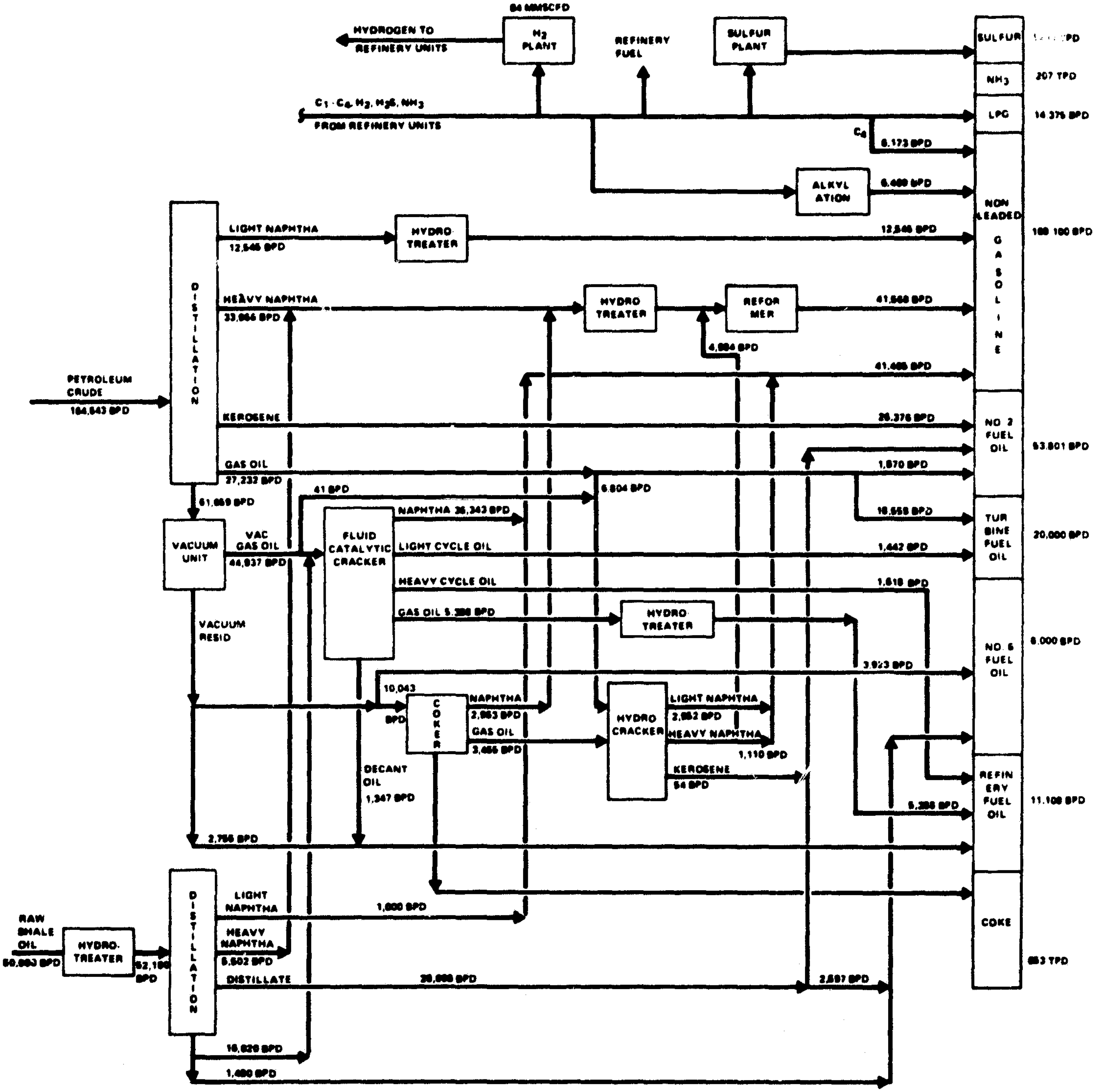


Figure 6-1 - Computer Data Output Diagram, Raw Shale Oil Plus Existing Refinery, Case 1

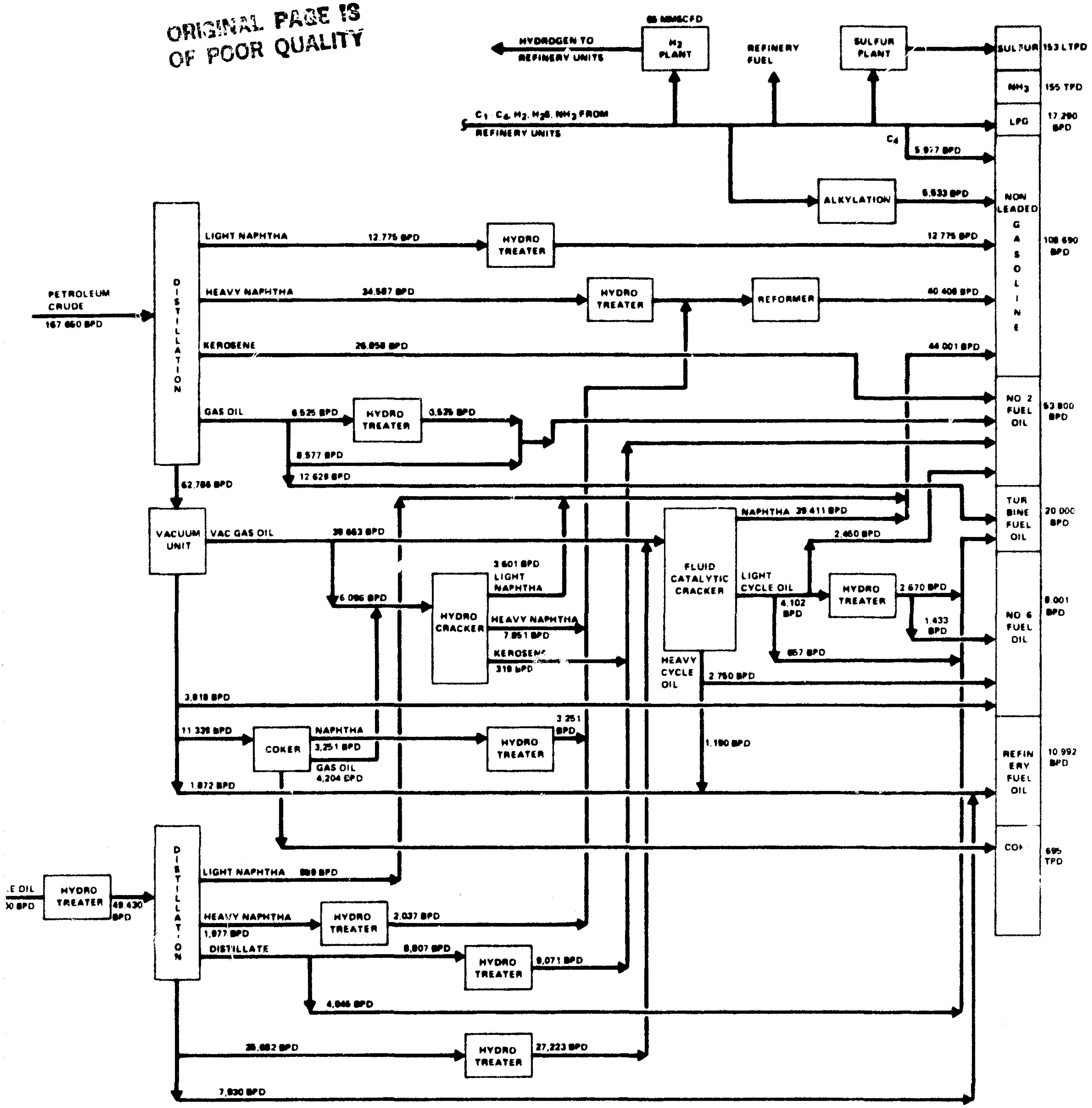
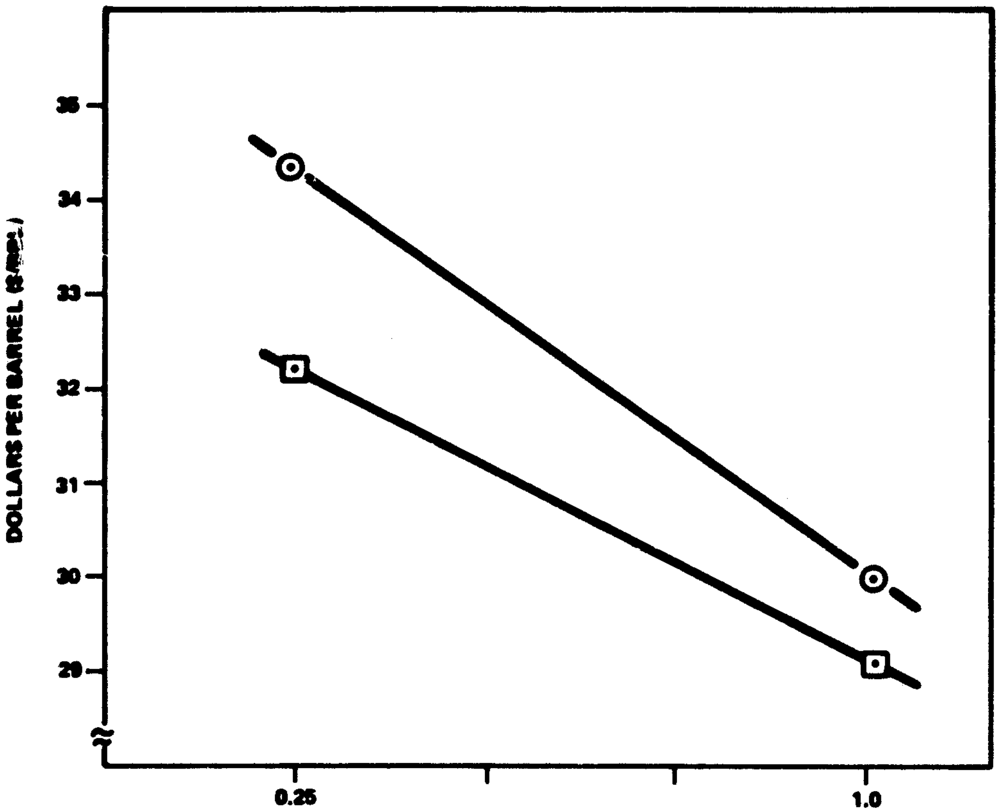


Figure 6-2 - Computer Data Output Diagram, Raw Shale Oil Plus Existing Refinery, Case 2

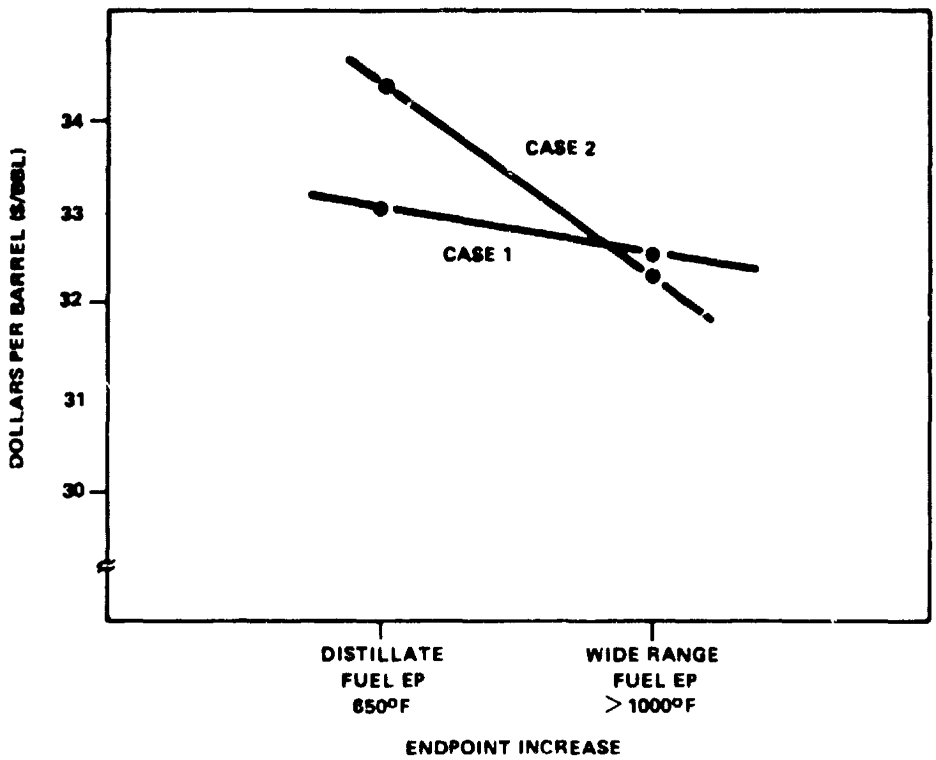


LEGEND:

⊙ TURBINE FUEL 1 (SPECIFICATION TF1, TABLE 6-1), FIGURE 6-2

⊡ TURBINE FUEL 3 (SPECIFICATION TF3, TABLE 6-1), FIGURE 6-2

Figure 6-3 - Effect of Varying the Nitrogen Specification of Turbine Fuel on Price, Raw Shale Oil Plus Existing Petroleum Refinery, Case 2

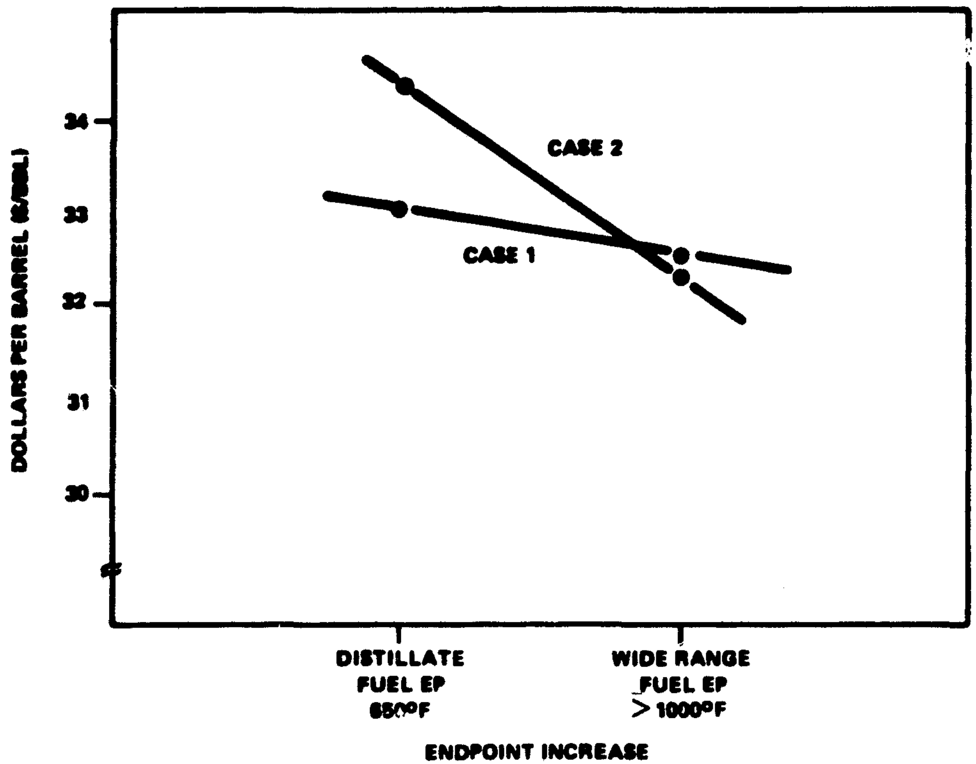


PROPERTIES OF TURBINE FUELS:

PROPERTY		TYPE OF FUEL							
		DISTILLATE CASE 1		WIDE RANGE FUEL CASE 1		DISTILLATE CASE 2		WIDE RANGE FUEL CASE 2	
		ACTUAL	SPECIFICATION TF1*	ACTUAL	SPECIFICATION TF3*	ACTUAL	SPECIFICATION TF1*	ACTUAL	SPECIFICATION TF3*
GRAVITY, °API	(MIN)	33.800	15.00	30.900	15.00	30.70	15.00	26.80	15.00
SULFUR, WT%	(MAX)	.036	0.70	0.700	0.70	0.70	0.70	0.70	0.70
NITROGEN, WT%	(MAX)	0.019	0.25	0.071	0.25	0.25	0.25	0.25	0.25
VISCOSITY (100°F), cst	(MAX)	5.100	5.80	5.200	160.00	4.30	5.80	5.60	160.00
FRACTION BOILING OVER 650°F, %		0.000	0.00	11.000	≤100.00	0.00	0.00	11.00	≤ 100.00

*SEE TABLE 6-1.

Figure 6-4 - Effect of Varying the Endpoint Specification of Turbine Fuel on Price, Raw Shale Oil Plus Existing Petroleum Refinery, Cases 1 and 2



PROPERTIES OF TURBINE FUELS:

PROPERTY		TYPE OF FUEL							
		DISTILLATE CASE 1		WIDE RANGE FUEL CASE 1		DISTILLATE CASE 2		WIDE RANGE FUEL CASE 2	
		ACTUAL	SPECIFICATION TF1[a]	ACTUAL	SPECIFICATION TF3[a]	ACTUAL	SPECIFICATION TF1[a]	ACTUAL	SPECIFICATION TF3[a]
GRAVITY, °API	(MIN)	33.800	15.00	30.900	15.00	30.70	15.00	26.80	15.00
SULFUR, WT%	(MAX)	.035	0.70	0.700	0.70	0.70	0.70	0.70	0.70
NITROGEN, WT%	(MAX)	0.019	0.25	0.071	0.25	0.25	0.25	0.25	0.25
VISCOSITY (100°F), cst	(MAX)	5.100	5.80	5.200	160.00	4.30	5.80	5.60	160.00
FRACTION BOILING OVER 650°F, %		0.000	0.00	11.000	≤100.00	0.00	0.00	11.00	≤100.00

[a]SEE TABLE 6-1.

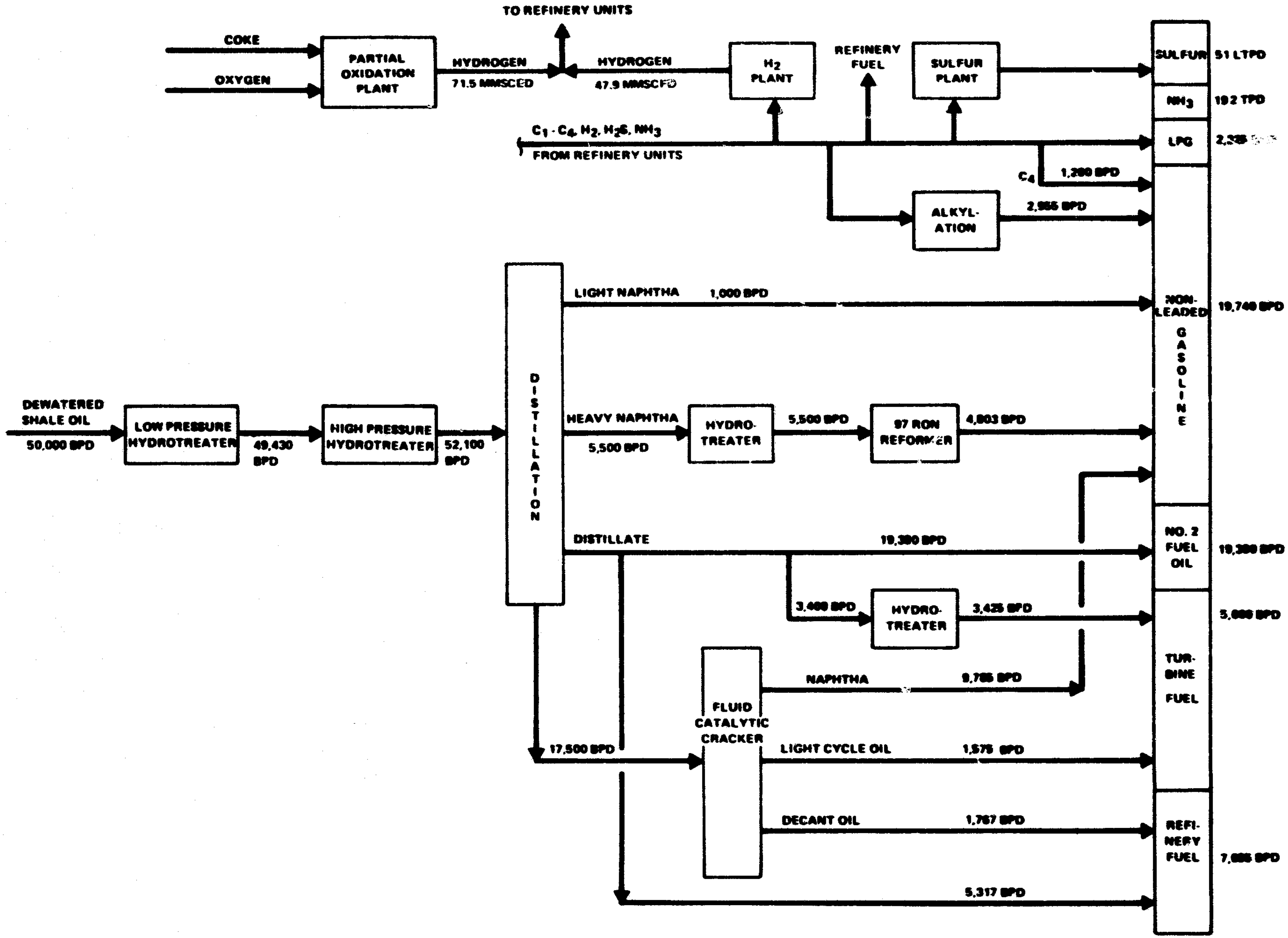


Figure 6-5 - Computer Output Data Diagram, New Shale Oil Refinery, Case 1

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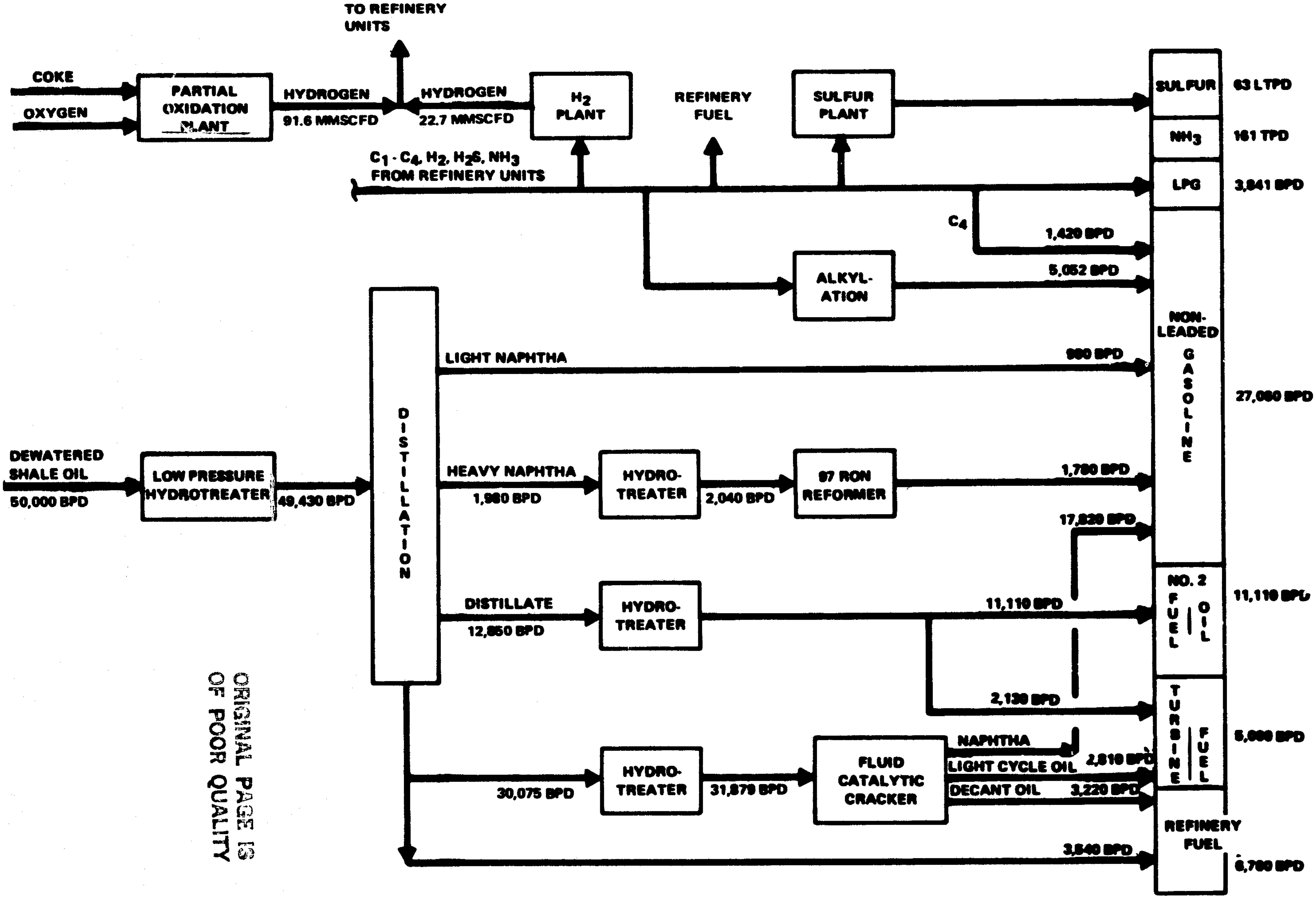


Figure 6-6 - Computer Output Data Diagram, New Shale Oil Refinery, Case 2

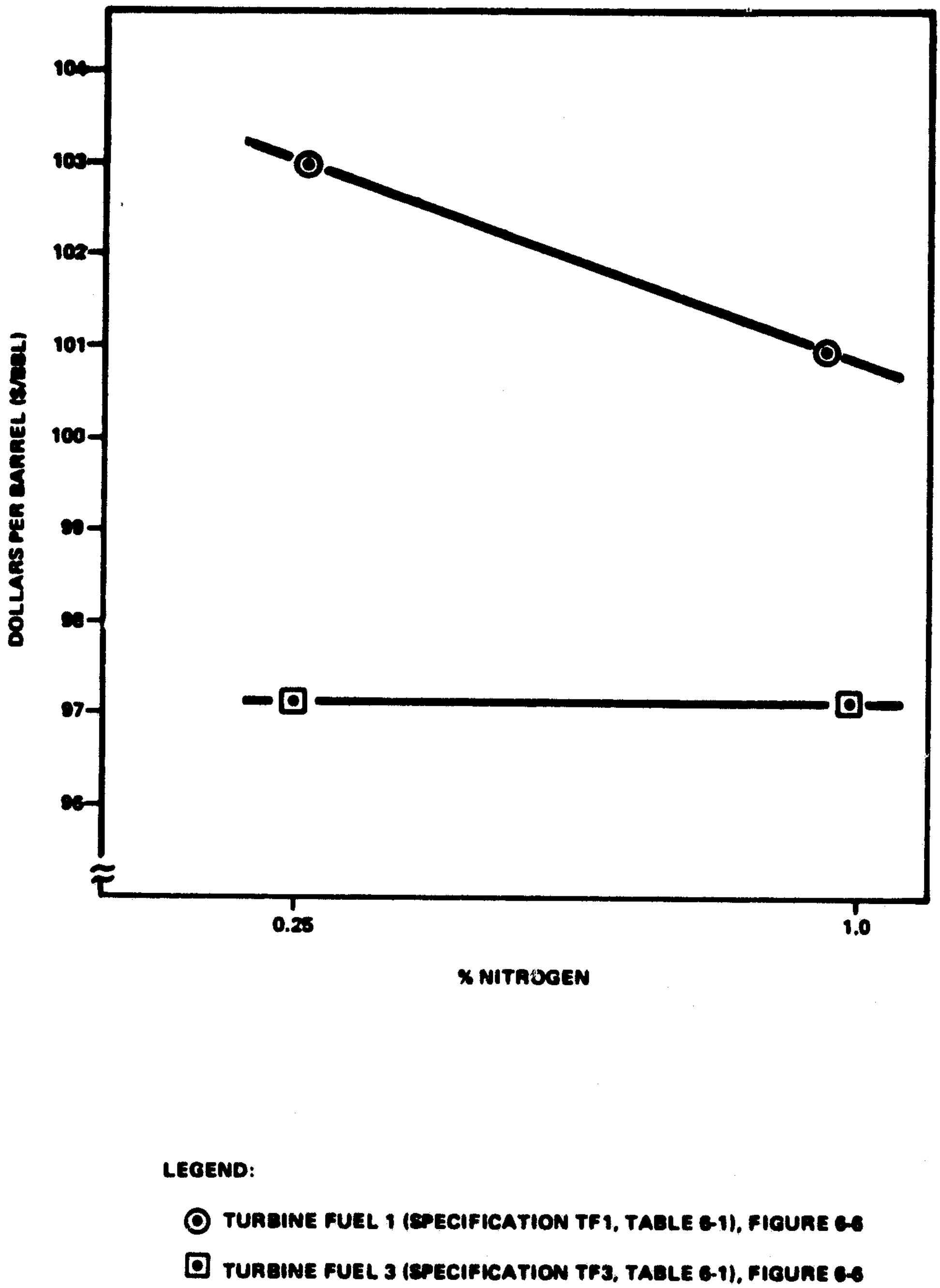
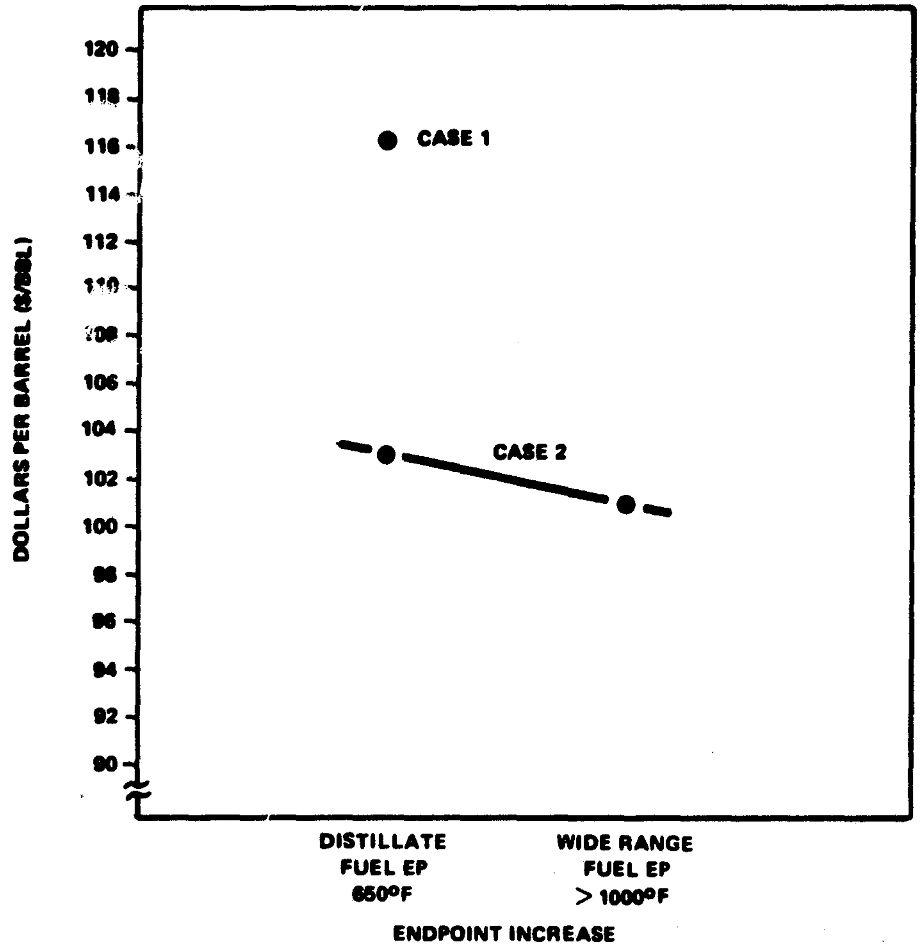


Figure 6-7 - Effect of Varying the Nitrogen Specification of Turbine Fuel on Price, New Shale Oil Refinery, Case 2



PROPERTIES OF TURBINE FUELS:

PROPERTY		TYPE OF FUEL					
		DISTILLATE CASE 1		DISTILLATE CASE 2		WIDE RANGE FUEL CASE 2	
		ACTUAL	SPECIFICATION TF1[a]	ACTUAL	SPECIFICATION TF1[a]	ACTUAL	SPECIFICATION TF3[a]
GRAVITY, °API	(MIN)	36.90	15.00	26.50	15.00	20.90	15.00
SULFUR, WT%	(MAX)	.0006	0.70	0.20	0.70	0.22	0.70
NITROGEN, WT%	(MAX)	0.05	0.25	0.25	0.25	0.25	0.25
VISCOSITY (100°F), cst	(MAX)	3.00	5.80	2.50	5.80	7.00	160.00
FRACTION BOILING OVER 650°F, %		0.00	0.00	0.00	0.00	50.00	≤ 100.00

[a]SEE TABLE 6-1.

Figure 6-8 - Effect of Varying the Endpoint Specification of Turbine Fuel on Price, New Shale Oil Refinery, Cases 1 and 2

6.5 H-COAL SYNFUEL UPGRADING

6.5.1 EXISTING REFINERY TO UPGRADE H-COAL

The refinery model combines the petroleum refinery with the H-Coal oil refinery by blending product streams to meet a given product slate. Additional process units are included where petroleum and H-Coal oil require separate treatment at different severity levels to meet product specifications. Based on a fixed feed of 50,000 BPD of H-Coal oil, and at a given production of gasoline, No. 2 fuel oil, No. 6 fuel oil, and turbine fuel, the program finds the most economical process route by reducing petroleum crude. To determine the impact of turbine fuel quality on the process economics, the process calculation for turbine fuel price was determined at different nitrogen levels and endpoint specifications for turbine fuel.

A. Refinery Linear Programming Output

The refinery configurations, resulting from linear programming calculation, show two major schemes. Case 1, Figure 6-9, shows the H-Coal oil being severely hydrotreated before fractionation and blending with petroleum products. In Case 2, Figure 6-10, more severe hydrotreating of fractions is applied, if necessary, after distillation of H-Coal oil. For both cases, refinery process calculations were completed for the following turbine fuel specifications:

	Case 1 (0.25% N)
TF1	distillate fuel
TF2	distillate fuel with higher viscosity limit
TF3	heavy fuel

High nitrogen fuels are not achievable in Case 1 because severe hydrotreating of whole H-Coal oil reduces nitrogen content below the turbine fuel specification.

Case 2

(.025% N)		(1.0% N)	
TF1	and	T11	(distillate fuel)
TF3	and	T13	(heavy fuel)

These specifications in Case 2 reflect the range of light and heavy fuel with low and high nitrogen.

B. Calculation of Turbine Fuel Prices

To determine the price of turbine fuel produced from a combined refinery consisting of petroleum and H-Coal oil feed, several basic operating conditions had to be defined, which are the same as applied to the shale oil cases. They are as follows:

(1) The amount of gasoline was held constant since the market for this fuel does not change, disregarding normal seasonal variations.

(2) The amount of No. 2 fuel should stay constant but can be reduced.

(3) 8,000 BPD of No. 6 fuel oil are produced.

(4) 20,000 BPD of turbine fuel have to be produced for the combined refinery cases.

(5) All product prices, except turbine fuel, are fixed. Thus turbine fuel price supports the profitability of the refinery expansion to meet a 15% discounted cash flow rate of return.

(6) The feed of H-Coal oil is fixed at 50,000 BPD, while crude oil feed can be reduced to meet a given product slate.

The required turbine fuel selling prices results of the different refinery configurations producing several grades of turbine fuel are shown in Table 6-14 and 6-15. Table 6-16 represents capital cost data for the combined refinery with severe hydrotreating before distillation, Case 1. Table 6-14 includes the calculated turbine fuel prices for turbine fuels TF1, TF2, and TF3 for Case 1.

Table 6-17 represents capital cost data for the combined refinery without mild hydrotreating before distillation, Case 2. Table 6-15 contains the calculated turbine fuel prices for turbine fuels TF1, T11, TF3, and T13 for Case 2.

The "capital recovery factor" described in Section 6.3.1B for shale oil is used in calculating turbine fuel prices for H-Coal plus the existing petroleum refinery as shown in Tables 6-14 and 6-15.

The product slates for Cases 1 and 2 are essentially unchanged for the several different turbine fuel quality specifications and are shown in Tables 6-18 and 6-19.

C. Evaluation of Turbine Fuel Prices Versus Turbine Fuel Quality

In the overall economics calculation, Tables 6-14, 6-15, the turbine fuel price reflects the change in refinery operation when turbine fuel specification is changed. Case 1, severe hydrotreating of H-Coal oil, TF1, TF2 and TF3 are blended showing no significant change of capital cost. Yet, the expansion of viscosity range and boiling range from TF1 to TF3 specification shows an increase in the difference between product value and feedcost which leads to a slight reduction of turbine fuel prices.

In Case 2, no hydrotreating before distillation, TF1 and TF3 specifications are applied, along with higher nitrogen level of 1 wt% (T11 and T13 respectively). The change of nitrogen limit shows a clear decrease of capital cost in both turbine fuel grades of approximately 6%. This is mainly the result of less mid-distillate hydrotreating in the H-Coal refinery.

The comparison of Cases 1 and 2 for both turbine fuel classifications TF1 and TF3, Figure 6-11, shows the influence of hydrotreating on turbine fuel prices. The severe hydrotreating in Case 1 lowers the nitrogen content of the blended products far below the specification limit without improving the economics of the whole refinery. The table in Figure 6-11 shows the actual properties of turbine fuel in comparison to the specification.

No direct conclusion can be drawn from changing the turbine fuel specification from distillate (TF1) to heavy fuel (TF3), because of the different production slate. Less No. 2 fuel oil was produced in the TF1 case, which gives different capital cost, feed and product values, but still shows the trend of decreasing turbine fuel price when the specifications are relaxed to higher viscosity and higher boiling point. The dominating restriction in this case was the sulfur limit of 0.7 wt% which determined the blending possibilities. Figure 6-12 shows the effect of higher nitrogen in turbine fuel on the price for both light and heavy turbine fuels. It also shows that the limit of 1 wt% nitrogen was not completely exploited, due to already low nitrogen content in the H-Coal oil fractions.

D. Thermal Efficiency

The thermal efficiencies of the H-Coal oil plus existing petroleum refining for Cases 1 and 2 and each of the turbine fuel specifications are shown in Table 6-20. The thermal efficiency for Case 1 turbine fuels TF1, TF2, and TF3 is about 90.0% for all fuels. The thermal efficiencies for Case 2 turbine fuels TF1, T11 and TF3, T13 is about 91.0% for all fuels.

E. Utilities

The utilities requirements shown in Table 6-21 are based on providing 1,250 psig steam for driving letdown turbines to provide power requirements and low level process steam. Fuel is provided from refinery fuel gas and fuel oil generated internally for firing heaters and boiler facilities. Cooling water, condensate, and sour water stripping facilities are also provided.

6.5.2 NEW H-COAL OIL REFINERY

Unlike the process calculation for the combined H-Coal oil plus petroleum refinery, the stand-alone H-Coal oil refinery is not given a product slate to meet, with the exception of 5,000 BPD of turbine fuel which has to be produced. LPG, gasoline, No. 2 and No. 6 fuel oil will be produced and blended to maximize the product value. With a fixed feed of 50,000 BPD H-Coal oil, the program finds the most economical process route, based on process yields and severity levels for the hydrotreatment of the different distillation fractions.

The refinery has to provide its own fuel for utility production. Hydrogen is produced from light gases from the refinery, but a unit for the partial oxidization of coke to hydrogen is included to provide the hydrogen shortfall which cannot be produced from refinery streams.

To determine the impact of turbine fuel quality on the process economics, the linear program model was allowed to blend to different turbine fuel specifications, as described in the combined refinery cases.

A. Resulting Refinery Linear Programming Configuration

The refinery configurations represent economical process routes for upgrading H-Coal oil in a new refinery when different hydrotreating methods are applied. The difference between the two configurations, Figures 6-13 and 6-14, is the degree of hydrotreating before and after distillation. In Figure 6-13, Case 1, severe hydrotreating at high pressure and low space velocity occurs to hydrodenitrify the whole H-Coal oil feed to a nitrogen level of about 50 ppm (wt). The result is an upgrading of whole H-Coal oil from an API of 30.5 to 40.4 degrees to a liquid suitable for further processing to petroleum specification products. The 650°F fraction results in an excellent feed for the FCC or hydrocracking process.

In Figure 6-14, Case 2, hydrotreating after distillation of individual fractions takes place at high pressures and low space velocity to

reduce the nitrogen to the level required to prevent poisoning and deactivation of the catalyst in subsequent processing units.

In none of the calculated Cases 1 and 2 was No. 6 fuel oil produced, due to the low viscosity of hydrocracked fuel oil and the small amount of high boiling fraction available for refinery fuel oil. In both of the calculated Cases 1 and 2, gasoline production was maximized as a means of increasing total product value. In Case 2, no No. 2 fuel oil was produced since the blendable fuel oil from the hydrocracking process was too heavy to meet No. 2 fuel oil specification. Also in Case 2, an excess of 1,444 BPD of turbine fuel was obtained since the fuel oil from hydrocracking could not be blended to No. 6 fuel oil due to viscosity restrictions.

In Figure 6-13, Case 1, no high nitrogen (1 wt%) turbine fuel (T11) was produced because of the severe hydrotreating of whole H-Coal oil which reduced the nitrogen level below 0.25 wt%. Thus, only two turbine fuels were obtained: a distillate turbine fuel (TF1) with a 650°F endpoint and less than 0.25 wt% nitrogen, and a distillate turbine fuel (TF2) with a 900°F endpoint and less than 0.25 wt% nitrogen.

In Figure 6-14, Case 2, the linear program was allowed to blend two different turbine fuel types: a distillate turbine fuel with a 650°F endpoint (TF1), and a turbine fuel like TF1 but with a wider viscosity range (TF2). However, it was found that the wider viscosity range allowed for TF2 was not obtainable due to the limits of Case 2, and only TF1 was obtained because extensive hydrotreating was performed to reduce nitrogen content to the point where it would not poison catalysts in downstream units.

B. Calculation of Turbine Fuel Prices

To determine a value for turbine fuel for the H-Coal oil refinery, the complete calculation was based on forcing the turbine fuel production of 5,000 BPD for Case 1 and for Case 2 at zero value. After deducting the daily capital recovery, operating cost and feed cost from the product value (excluding turbine fuel), a revenue margin was left which had

to be supported by the turbine fuel price. This price represents the required revenue of turbine fuel to an H-Coal oil refinery forced to produce 5,000 BPD for Case 1 and a resulting 6,444 BPD of turbine fuel for Case 2. The turbine fuel price is based on selling all other products with petroleum specifications at the market prices prevailing for comparable petroleum products.

Table 6-22 presents capacity and capital cost data for the new H-Coal refinery for Case 1 which includes TF1 and TF2, and Case 2 which includes TF1 only. Table 6-23 includes the calculated tubine fuel required prices for TF1, TF2 for Cases 1 and 2 for the H-Coal oil refinery.

C. Evaluation of Turbine Fuel Prices Versus Turbine Fuel Quality

The evaluation of the turbine fuel required price calculations, as shown in Table 6-23, indicates the key factors that affect prices are as follows:

(1) The data in Table 6-23, severe hydrotreating before distillation, Case 1, TF1, TF2 indicates a high turbine fuel price is required to support the 15% discounted cash flow profit level of the new H-Coal oil refinery. This price range is about \$114-\$121, or about three times the combined H-Coal plus petroleum refinery turbine fuel price.

The factor that most affects this price difference is the high capital investment cost for the new H-Coal refinery of \$8,000 per daily barrel. The comparable 200,000 BPD petroleum refinery cost is \$3,000 per daily barrel.

(2) The data in Table 6-23, Case 2, severe hydrotreating after distillation, TF1, indicates a lower fixed capital investment cost for the new H-Coal oil refinery as compared with Case 1, a higher product value, and an

increase of 1,444 BPD in turbine fuel obtained. This results in a much lower turbine fuel price of about $67 per barrel. Disregarding the increased fuel obtained, the major effect on the turbine fuel required price is the change in capital investment cost caused by deleting the high pressure hydrotreater and adding more hydrocracker capacity. If only 5,000 BPD of turbine fuel is produced, instead of 6,444 BPD, the turbine fuel price would be about $85 per barrel.

An evaluation of price versus endpoint specification for turbine fuels is shown in Figure 6-15 for Cases 1 and 2. These curves are a plot of the calculated turbine fuel required prices versus two distillate type turbine fuels with different endpoints. The properties of these distillate turbine fuels are shown in the table below Figure 6-15.

To produce the wide range turbine fuel, about 50% of the blended fuel is in a boiling range over 650°F which results in a lower gravity and slightly higher viscosity of the product.

D. Thermal Efficiency

The thermal efficiencies of the new H-coal oil refinery for Cases 1 and 2 and each of the turbine fuel specifications are shown in Table 6-24. The thermal efficiencies for Case 1 turbine fuels TF1 and TF2 are 85.0% for both fuels. The thermal efficiency for Case 2 turbine fuel TF1 is 86.0%.

E. Utilities

The utilities requirement shown in Table 6-25 are based on providing a 1,250 psig steam plant for driving letdown turbines to provide power requirement and lower pressure process steam. Fuel is provided from

refinery fuel gas and fuel oil generated internally for firing heaters and boilers. Cooling water circulation, condensate recovery, and sour water stripping facilities are also provided.

Table 6-14 - Turbine Fuel Selling Prices, Petroleum Crude Plus H-Coal Oil Refinery, 20,000 BPD Turbine Fuel Produced, $ per Day

Item	Case i Severe Hydrotreating 0.25% Nitrogen		
	Distillate Fuel 650°F Endpoint (TF1)	Distillate Fuel Below 1000°F Endpoint (TF2)	Heavy Fuel Above 1000°F Endpoint (TF3)
Fixed Capital Investment for Additional Process Units	149,100,000	149,200,000	149,000,000
Add: Offsite Facilities $\frac{\text{FCI Add'tl Proc. Units} \times 0.30}{0.70}$	63,900,000	63,900,000	63,900,000
Royalties and Catalyst	4,330,000	4,130,000	4,120,000
Total Additional Capital Investment	217,330,000	217,230,000	217,020,000
Daily Capital Recovery (Total Add'tl Cap x 0.0010606*)	230,500	230,394	230,171
Add: Feed Cost	6,578,510	6,508,120	6,470,190
Operating Cost	54,080	54,020	54,050
Total Daily Required Revenue	6,863,090	6,792,534	6,754,411
Deduct: Product Values (Exclusive of Turbine Fuel)	6,697,310	6,629,100	6,598,450
Revenue Margin	165,780	163,434	155,961
Add: Operating Margin of Existing Refinery Before Addition of Synfuel Upgrading	702,204	702,204	702,204
Turbine Fuel Required Daily Revenue	867,984	865,638	858,165
Minimum Selling Price Per Barrel Turbine Fuel	43.40	43.28	42.91

* Capital Recovery Factor, stream day basis: $\frac{0.35}{\text{330 operating days p.a.}} = 0.0010606$

Table 6-15 - Turbine Fuel Selling Prices, Petroleum Crude Plus H-Coal Oil Refinery, 20,000 BPD Turbine Fuel Produced, $ per Day

Item	Case 2 TF1	T11	TF3	T13
Fixed Capital Investment for Additional Process Units	104,700,000	77,900,000	110,400,000	83,100,000
Add: Offsite Facilities $\frac{\text{FCI Add'tl Proc. Units} \times 0.30}{0.70}$	44,900,000	33,400,000	47,300,000	35,600,000
Royalties and Catalyst	2,176,000	1,388,000	2,659,000	1,512,000
Total Additional Capital Investment	151,776,000	112,688,000	160,959,000	120,212,000
Daily Capital Recovery (Total Add'tl Cap x 0.0010606*)	160,974	119,517	170,713	127,497
Add: Feed Cost	6,382,450	6,459,420	6,616,230	6,534,900
Operating Cost	51,960	51,650	52,755	51,814
Total Daily Required Revenue	6,595,384	6,630,587	6,839,698	6,714,211
Deduct: Product Values (Exclusive of Turbine Fuel)	6,518,880	6,600,730	6,765,830	6,693,400
Revenue Margin	76,504	29,857	73,868	20,811
Add: Operating Margin of Existing Refinery Before Addition of Synfuel Upgrading	702,204	702,204	702,204	702,204
Turbine Fuel Required Daily Revenue	778,708	732,061	776,072	732,015
Minimum Selling Price Per Barrel Turbine Fuel	38.94	36.60	38.80	36.15

* Capital Recovery Factor, stream day basis: $\frac{0.35}{330 \text{ operating days p.a.}} = 0.0010606$

Table 6-16 - Capacity and Capital Cost Data, Petroleum Plus H-Coal Oil Refinery, Case 1

Process Unit	Unit Capacities, BPD				Fixed Capital Investment ($ Million)		
	Existing Refinery	Equipment Additions TF1	Equipment Additions TF2	Equipment Additions TF3	TF1	TF2	TF3
Crude Unit	200,000	—	—	—	—	—	—
Vacuum Distillation	75,000	—	—	—	—	—	—
Fluid Catalytic Cracker	50,000	—	—	—	—	—	—
Hydrocracker	10,300	—	—	—	—	—	—
Coker	12,500	—	—	—	—	—	—
Naphtha Hydrotreater	61,000	3,588	2,670	1,980	2.1	1.7	1.4
Atm Gas Oil Hydrotreater	22,000	—	—	—	—	—	—
Reformer	49,000	3,060	2,160	1,980	7.1	5.6	5.3
Alkylation	8,0000	—	—	—	—	—	—
H-Coal Oil H.P. Hydrotreater	—	50,000	50,000	50,000	114.9	114.9	114.9
H-Coal Oil Distillation	—	51,800	51,800	51,800	14.0	14.0	14.0
Hydrogen Plant, million SCFD	—	3.66	4.03	5.88	3.3	3.5	4.6
Sulfur Recovery Plant, long ton/day	135	—	—	—	—	—	—
Ammonia recovery from Waste Water, ton/day NH_3	17	33	33	33	2.6	2.6	2.6
Sour Water Stripper, M lb/day	5,300	959	997	918	0.7	0.8	0.7
Cooling Water System, M gal/day	196,000	—	—	—	—	—	—
Steam/Power Plant, M lb/day, 1250 psig steam	15,100	444	654	571	4.5	6.1	5.5
Total Additional FCI					149.1	149.2	149.0

Table 6-17 - Capacity and Capital Cost Data, Petroleum Plus H-Coal Oil Refinery, Case 2

Process Unit	Unit Capacity, BPD					Fixed Capital Investment ($ Million)			
	Existing Refinery	Equipment Additions							
		TF1	T11	TF3	T13	TF1	T11	TF3	T13
Crude Unit	200,000	—	—	—	—	—	—	—	—
Vacuum Distillation	75,000	—	—	—	—	—	—	—	—
Fluid Catalytic Cracker	50,000	—	—	—	—	—	—	—	
Hydrocracker	10,300	—	—	—	—	—	—	—	—
Coker	12,500	—	—	—	—	—	—	—	—
Naphtha Hydrotreater	61,000	—	—	—	—	—	—	—	—
Atm Gas Oil Hydrotreater	22,000	—	—	—	—	—	—	—	—
Reformer	49,000	817	—	2,730	—	2.8	—	6.6	—
Alkylation	8,000	—	—	—	—	—	—	—	—
H-Coal Oil Distillation	—	50,000	50,000	50,000	45,960	13.5	13.5	13.5	13.5
H-Coal Oil Naphtha Hydrotreater	—	18,500	18,500	18,500	18,500	49.5	49.5	49.5	49.5
H-Coal Oil Distillate Hydrotreater	—	11,357	1,010	12,336	3,614	36.9	8.6	38.8	18.6
H-Coal Oil Heavy Gas Oil Hydrotreater	—	—	695	—	—	—	5.0	—	—
Hydrogen Plant, million SCFD	—	—	—	—	—	—	—	—	—
Sulfur Recovery Plant, Long ton/day	135	—	—	—	—	—	—	—	—
Ammonia Recovery from Waste Water, ton/day NH_3	17	21	11	22	14	2.0	1.3	2.0	1.5
Sour Water Stripper, M lb/day	5,300	—	—	—	—	—	—	—	—
Cooling Water System, M gal/day	196,000	—	—	—	—	—	—	—	—
Steam/Power Plant, M lb/day, 1250 psig steam	15,100	—	—	—	—	—	—	—	—
Total Additional FCI						104.7	77.9	110.4	83.1

Table 6-18 - Product Slate, H-Coal Oil Plus Petrolelum Refinery, Case 1

Item	Rate	TF1	TF2	TF3
Feed				
Petroleum Crude	M BPD	169.95	163.6	162.34
H-Coal Oil	M BPD	50.0	50.0	50.0
Products				
LPG	M BPD	17.3	14.6	13.5
Gasoline	"	108.1	108.1	108.1
No. 2 fuel oil	"	53.8	53.8	53.8
No. 6 fuel oil	"	8.0	8.0	8.0
Turbine fuel	"	20.0	20.0	20.0
Sulfur	M LTPD	0.101	0.101	0.101
Ammonia	M TPD	0.048	0.048	0.048
Coke	M TPD	0.706	0.650	0.575

Table 6-19 - Product Slate, H-Coal Oil Plus Petroleum Refinery, Case 2

		Turbine Fuels			
Item	Rate	TF1	T11	TF3	T13
Feed					
Petroleum Crude	M BPD	159.42	162.0	167.2	164.5
H-Coal Oil	M BPD	50.0	50.0	50.0	50.0
Products					
LPG	M BPD	17.9	19.5	20.2	19.0
Gasoline	"	108.1	108.1	108.1	108.1
No. 2 fuel oil	"	47.8	49.1	53.8	53.8
No. 6 fuel oil	"	8.0	8.0	8.0	8.0
Turbine fuel	"	20.0	20.0	20.0	20.0
Sulfur	M LTPD	0.109	0.117	0.113	0.115
Ammonia	M LTPD	0.038	0.028	0.039	0.031
Coke	M TPD	0.676	0.689	0.710	0.700

Table 6-20 - Thermal Efficiencies of H-Coal Oil Plus Existing Petroleum Refinery

	Million Btu/D						
	Case 1			Case 2			
Item	TF1	TF2	TF3	TF1	T11	TF3	T13
Total Heating Value Feed	1245.7	1232.2	1224.9	1208.0	1222.8	1253.0	1237.3
Total Heating Value Products	1124.6	1111.6	1104.1	1096.5	1110.5	1140.4	1124.8
Thermal Efficiency, %	90.3	90.2	90.1	90.7	90.8	91.0	90.9

Table 6-21 - Total Utilities Requirement, H-Coal Oil Plus Existing Petroleum Refinery (Computer Output)

	Usage Rate	
Unit	Case 1	Case 2
Sour Water stripping	6259 M lb/D	5193 M lb/D
Cooling water circulation	186 MM gal/D	182 MM gal/D
Power generation	1277 M kWh/D	1233 M kWh/D
Fuel consumption	72.7 MMM Btu/D	74.1 MMM Btu/D

Table 6-22 - Capacity and Capital Cost Data, New H-Coal Oil Refineries, Case 1 and Case 2

Process Units	Unit Capacity, BPD Case 1 (0.25%N) TF1	Unit Capacity, BPD Case 1 (0.25%N) TF2	Unit Capacity, BPD Case 2 (0.25%N) TF1	Fixed Capital Investment ($ Million) Case 1 TF1	Fixed Capital Investment ($ Million) Case 1 TF2	Fixed Capital Investment ($ Million) Case 2 TF1
High Pressure Hydrotreater	50,000	50,000		114.9	114.9	
Distillation	51,800	51,800	50,000	14.0	14.0	13.5
Naphtha Hydrotreater	14,504	14,504	18,500	42.8	42.8	49.5
Hydrocracker	3,972	—	27,500	25.4	—	81.1
Reformer	16,124	14,504	26,516	22.8	21.2	32.3
Hydrogen Plant, million SCFD	37.9	34.8	60.2	17.0	16.0	23.5
Sulfur Recovery Plant, long ton/day	10	10	10	0.9	0.9	0.9
Ammonia Recovery from Waste Water, ton/day NH_3	34	34	32	2.6	2.6	2.5
Sour Water Stripper, M lb/day	2,226	1,957	2,615	1.4	1.3	1.6
Cooling Water System, M gal/day	54,789	49,634	83,120	2.2	2.0	3.0
Steam/Power Plant, M lb/day, 1250 psig steam	3,482	2,851	4,994	23.2	19.8	30.9
Total Fixed Capital Investment				267.2	235.5	238.8

Table 6-23 - Turbine Fuel Selling Prices, New H-Coal Oil Refinery, 5,000 BPD Turbine Fuel Produced, $ Per Day

Item	Case 1 TF1 (0.25% N) 650°F EP	Case 1 TF2 (0.25% N) 1000°F EP	Case 2[a] TF1 (0.25% N) 650°F EP
Fixed Capital Investment for Process Units	267,200,000	235,500,000	238,800,000
Add: Offsite Facilities $\frac{\text{FCI Add'tl Proc. Units} \times 0.30}{0.70}$	114,514,000	100,929,000	102,343,000
Royalties and Catalyst	8,682,000	7,703,000	13,407,000
Total Capital Investment	390,396,000	344,132,000	354,550,000
Daily Capital Amortization (Total Add'tl Cap x 0.0010606)	414,054	364,986	376,036
Add: Feed Cost	1,600,000	1,600,000	1,600,000
Operating Cost	22,623	20,512	23,042
Total Daily Cost	2,036,677	1,985,498	1,999,078
Deduct: Product Values (Exclusive of Turbine Fuel)	1,429,700	1,416,000	1,569,600
Turbine Fuel Required Daily Revenue	606,977	569,498	429,478
Minimum Selling Price Per Barrel Turbine Fuel	121.40	113.90	66.65

[a] Based on production of 6,444 BPD Turbine Fuel.

Table 6-24 - Thermal Efficiencies of New H-Coal Oil Refinery

Item	Million Btu/D Case 1 TF1	Case 1 TF2	Case 2 TF1
Total Heating Value Feed	287.2	287.2	287.2
Total Heating Value Product	243.5	244.8	247.0
Thermal Efficiency, %	84.8	85.2	86.0

Table 6-25 - New H-Coal Oil Refinery Total Utilities Requirement
(Computer output)

Unit	Usage Rate Case 1	Case 2
Sour Water Stripping	2091 M lb/D (17.4 gpm)	2615 M lb/D (2.8 gpm)
Cooling Water Circulation	52.2 M gal/D (36 gpm)	83.1 M gal/D (58 gpm)
Power Generation	349.7 M kWh/D (14,570 kW)	610.6 M kWh/D (25,440 kW)
Fuel Consumption	11.4 MMM Btu/D	15.4 MMM Btu/D

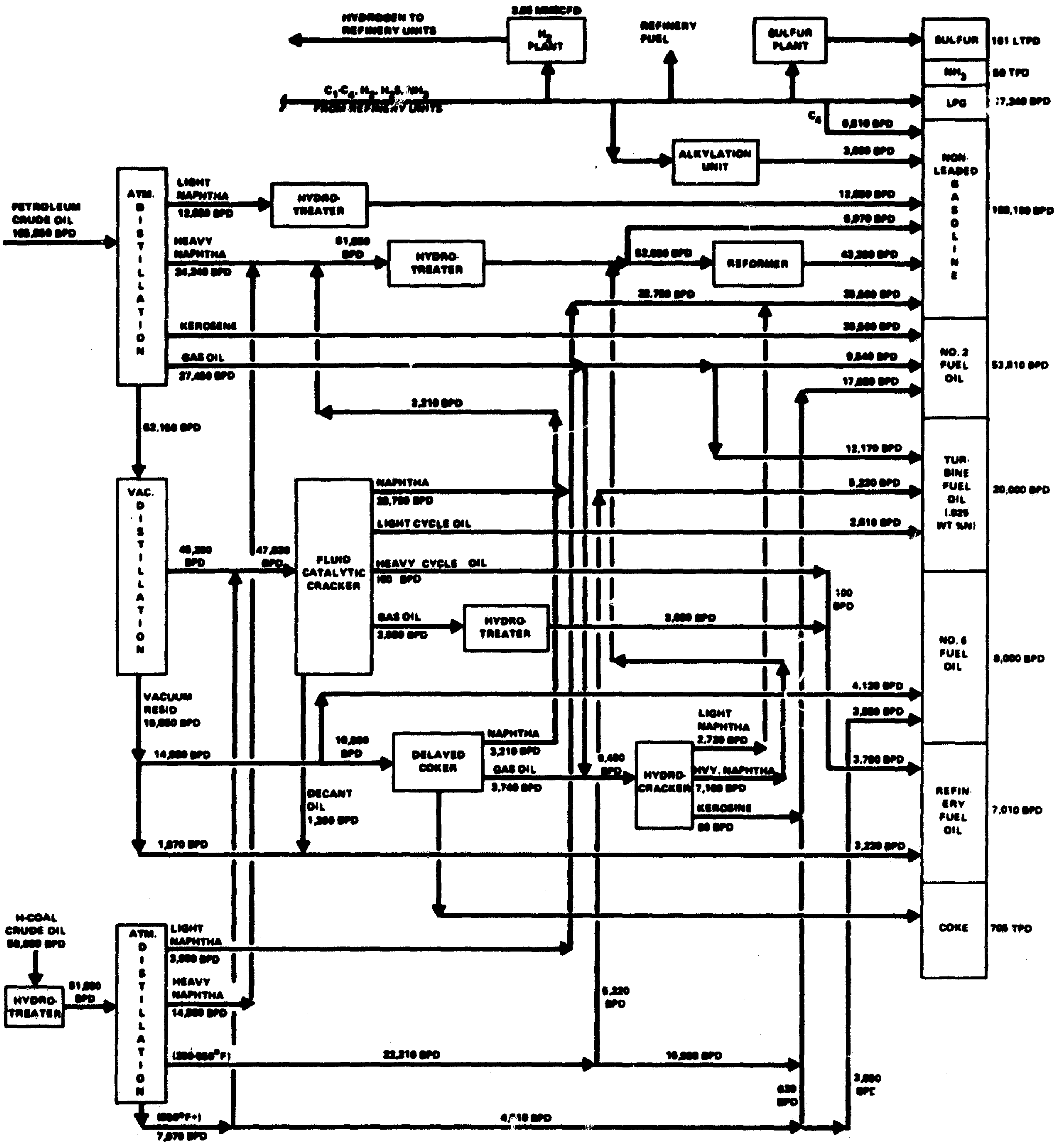


Figure 6-9 - Computer Output Data Diagram, H-Coal Plus Existing Petroleum Refinery, Case 1

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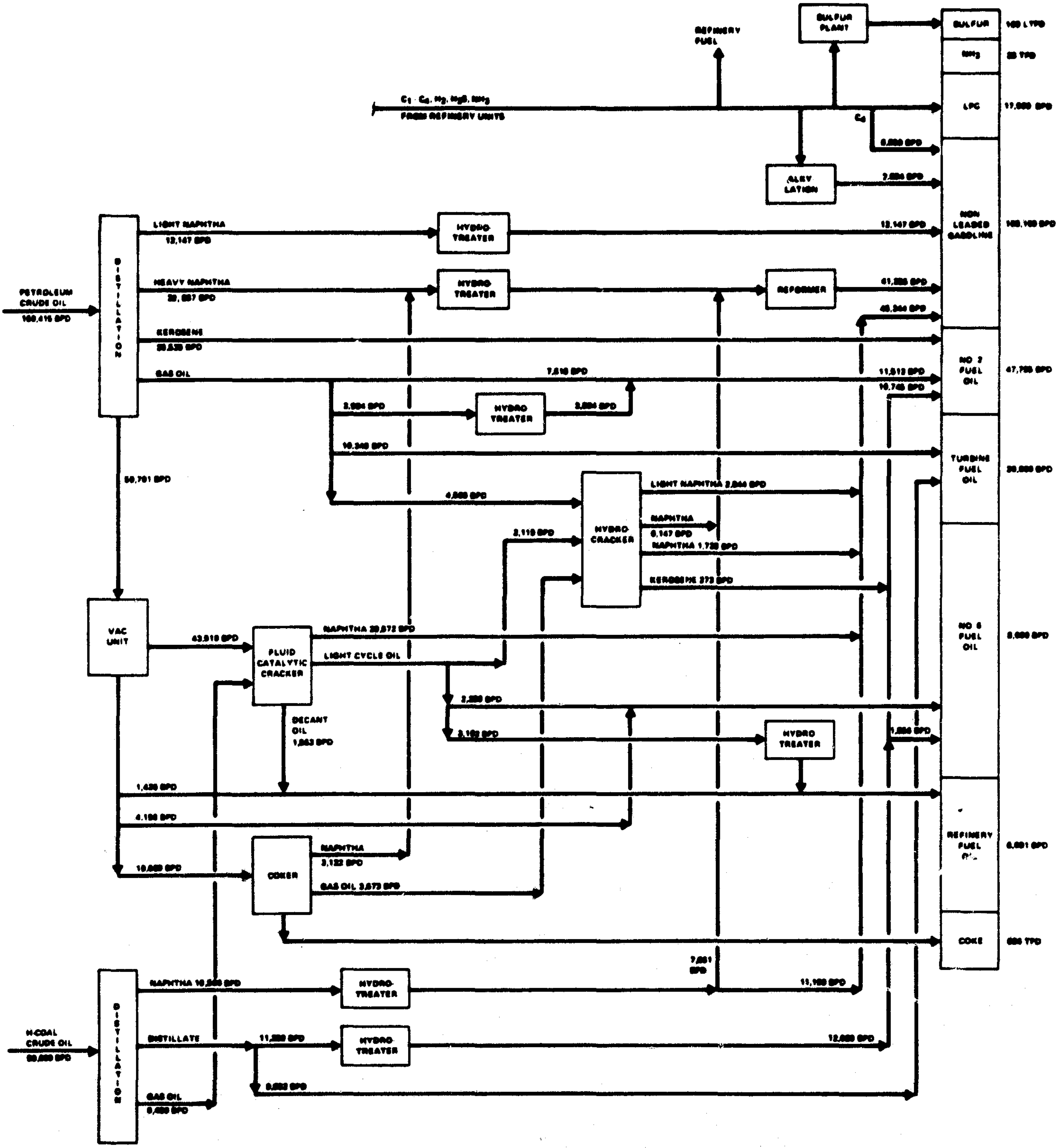

Figure 6-10 - Computer Output Data Diagram, H-Coal Oil Plus Existing Refinery, Case 2

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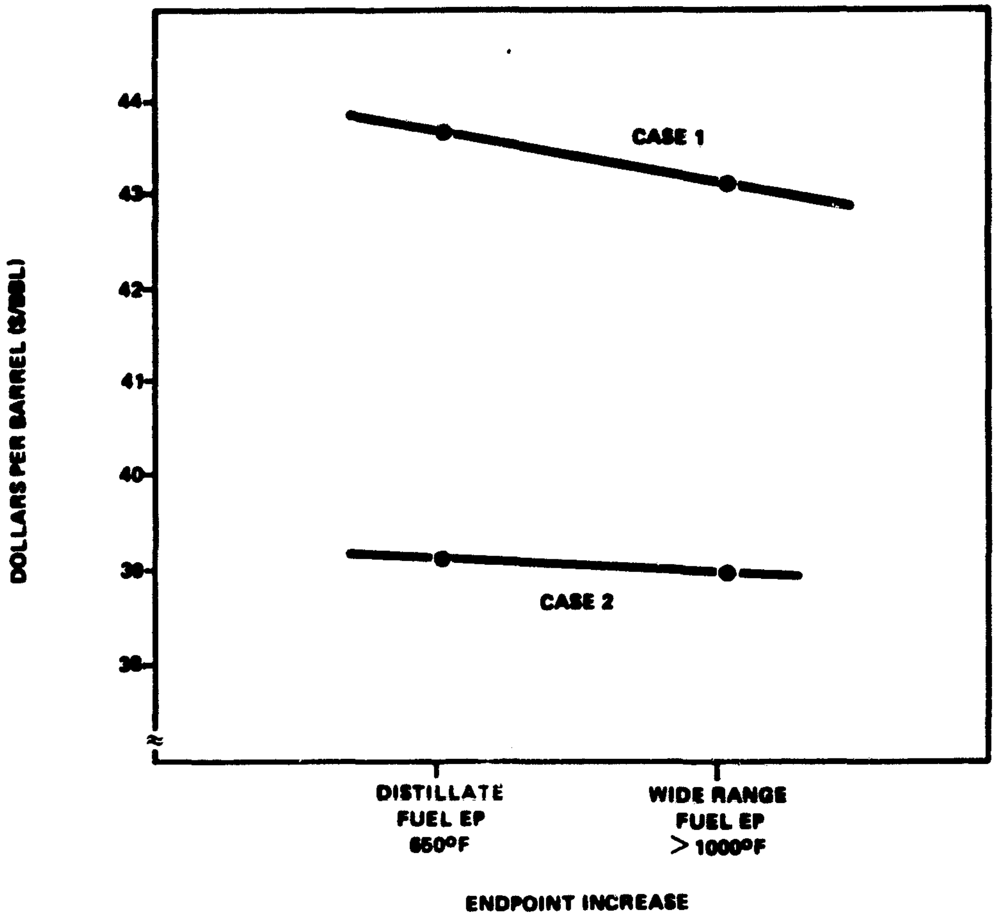
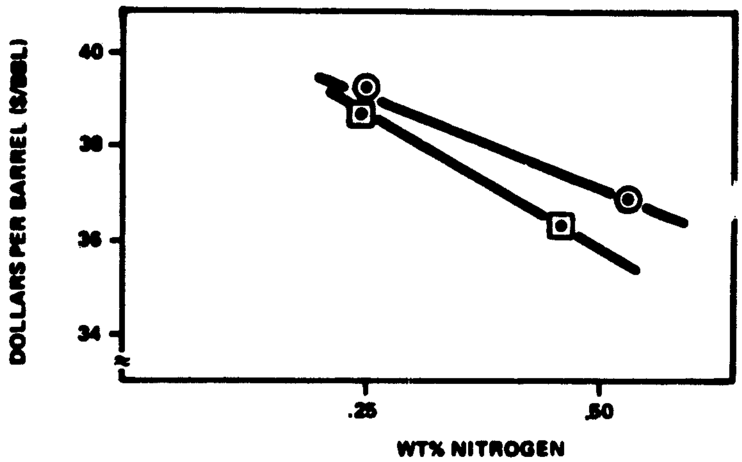


PROPERTIES OF TURBINE FUELS:

PROPERTY		TYPE OF FUEL							
		DISTILLATE CASE 1		WIDE RANGE FUEL CASE 1		DISTILLATE CASE 2		WIDE RANGE FUEL CASE 2	
		ACTUAL	SPECIFICATION TF1*	ACTUAL	SPECIFICATION TF3*	ACTUAL	SPECIFICATION TF1*	ACTUAL	SPECIFICATION TF3*
GRAVITY, °API	(MIN)	30.600	15.00	27.800	15.00	26.80	15.00	25.10	15.00
SULFUR WT%	(MAX)	0.700	0.70	0.700	0.70	0.40	0.70	0.70	0.70
NITROGEN, WT%	(MAX)	0.057	0.25	0.047	0.25	0.25	0.25	0.25	0.25
VISCOSITY (100°F), cst	(MAX)	4.100	5.80	7.200	160.00	4.00	5.80	5.80	160.00
FRACTION BOILING OVER 650°F, %		0.000	0.00	13.000	≤ 100.00	0.00	0.00	9.00	≤ 100.00

*SEE TABLE 6-1.

Figure 6-11 - Effect of Varying the Endpoint Specification of Turbine Fuel on Price, H-Coal Oil Plus Existing Petroleum Refinery, Cases 1 and 2



LEGEND:

◉ TURBINE FUEL 1 (SPECIFICATION TF1, TABLE 6-1), FIGURE 6-10

▣ TURBINE FUEL 3 (SPECIFICATION TF3, TABLE 6-1), FIGURE 6-10

Figure 6-12 - Effect of Varying the Nitrogen Specification of Turbine Fuel on Price, H-Coal Oil Plus Existing Petroleum Refinery, Case 2

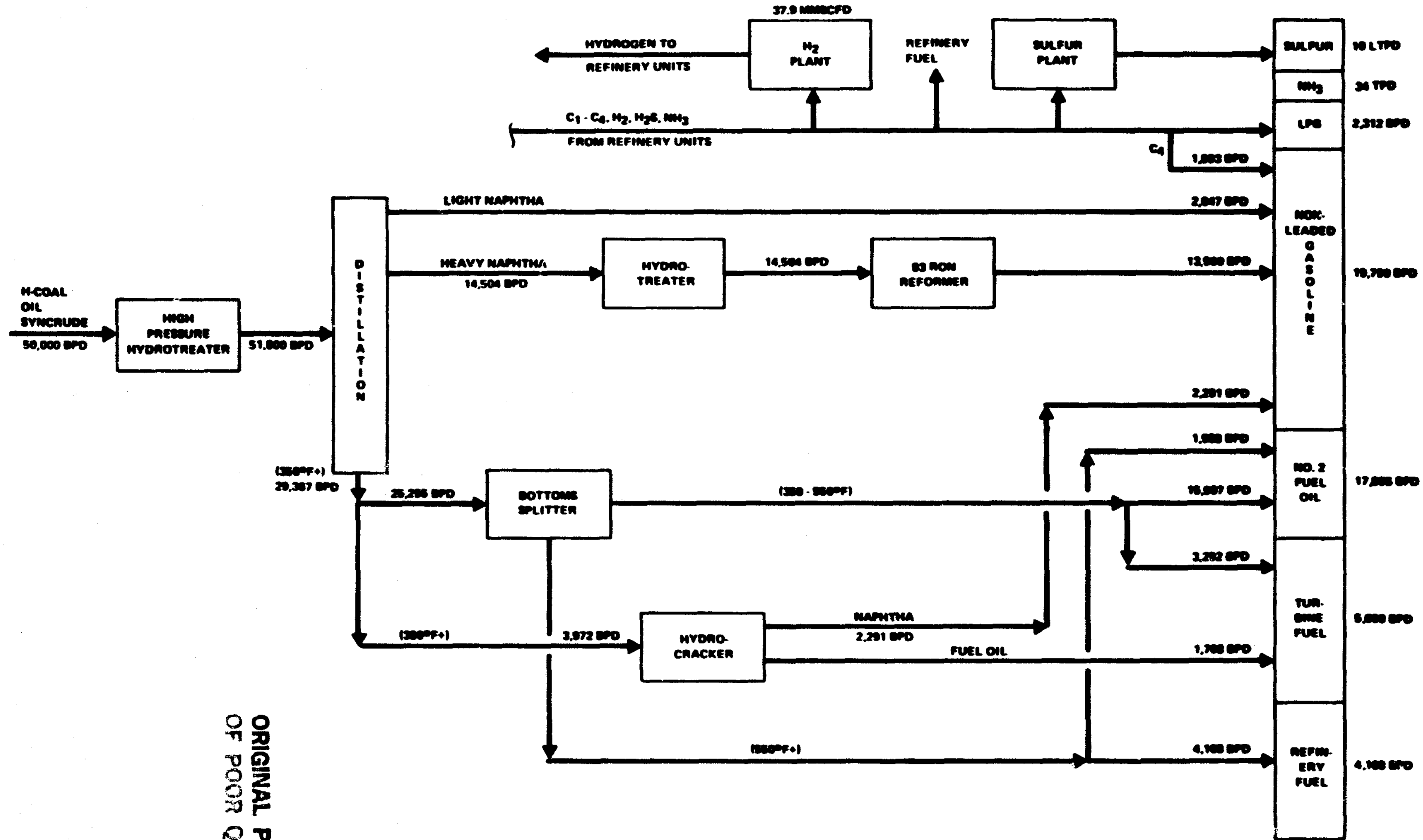


Figure 6-13 - Computer Output Data Diagram,
New H-Coal Oil Refinery, Case 1

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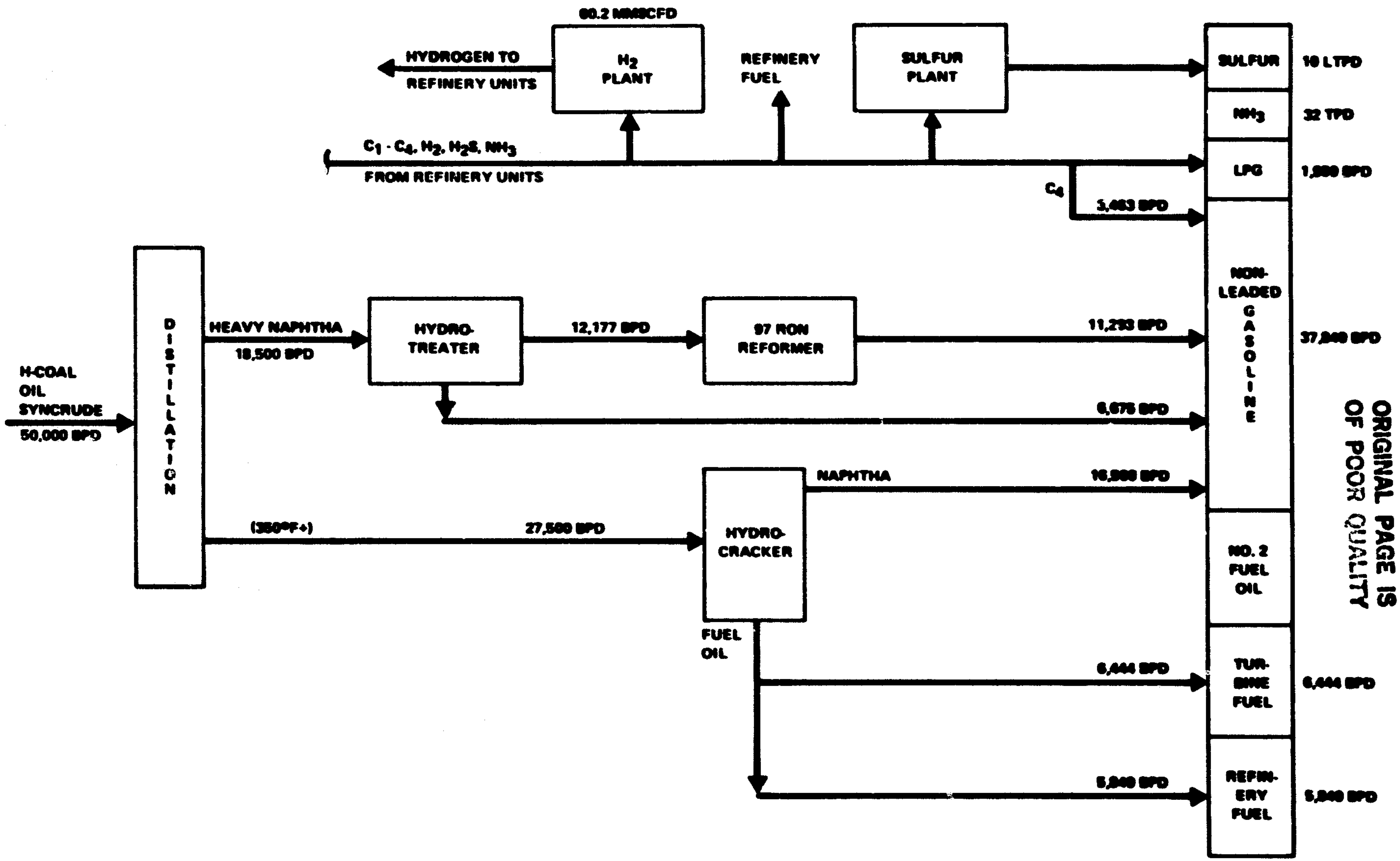
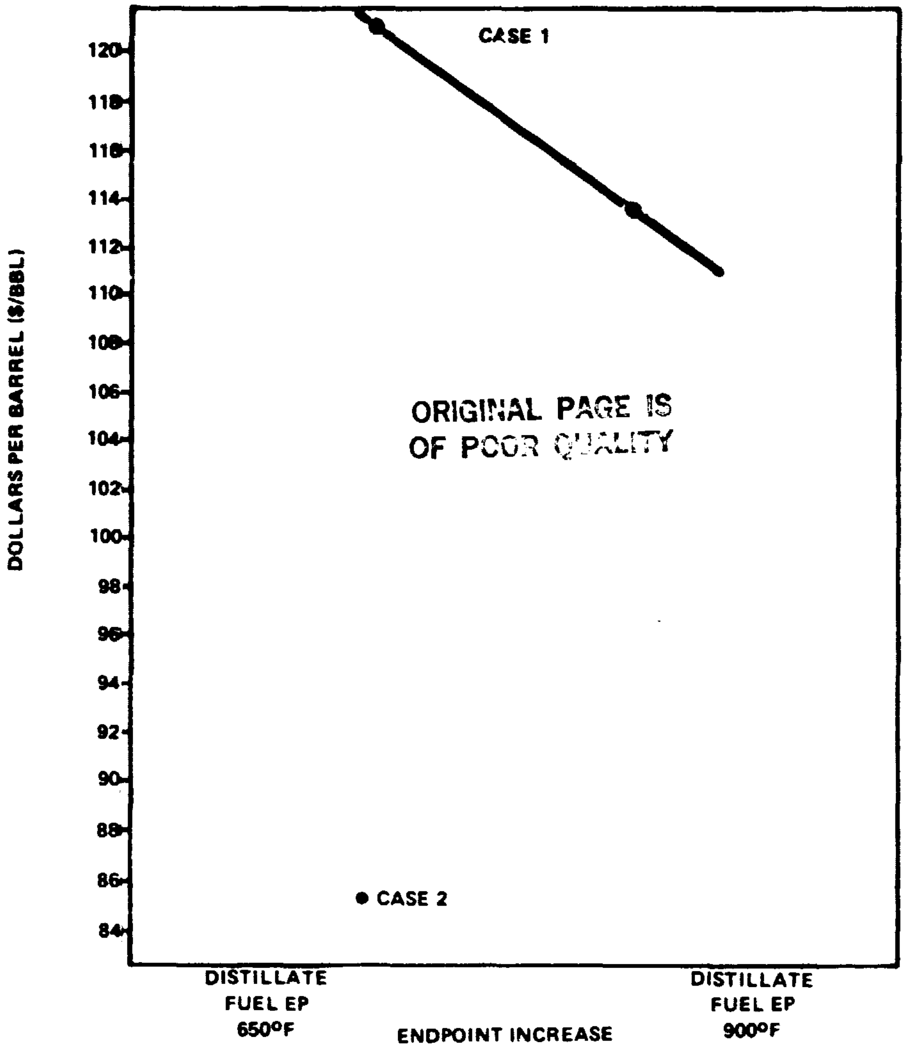


Figure 6-14 - Computer Output Data Diagram, New H-Coal Oil Refinery, Case 2



PROPERTIES OF TURBINE FUELS:

PROPERTY		TYPE OF FUEL					
		DISTILLATE CASE 1		DISTILLATE FUEL CASE 1		DISTILLATE CASE 2	
		ACTUAL	SPECIFICATION TF1*	ACTUAL	SPECIFICATION TF2*	ACTUAL	SPECIFICATION TF1*
GRAVITY, °API	(MIN)	29.000	15.00	23.300	15.00	24.300	15.00
SULFUR, WT%	(MAX)	.0001	0.70	.004	0.70	.001	0.70
NITROGEN, WT%	(MAX)	.002	0.25	.006	0.25	.002	0.25
VISCOSITY (100°F), cst	(MAX)	3.400	5.80	3.500	30.00	5.800	5.80
FRACTION BOILING OVER 650°F, %		0.000	0.00	50.000	≤ 100.00	0.000	0.00

*SEE TABLE 6-1.

Figure 6-15 - Effect of Varying the Endpoint Specification of Turbine Fuel on Price, New H-Coal Oil Refinery, Cases 1 and 2

6.6 SRC-II SYNFUEL UPGRADING

6.6.1 EXISTING REFINERY TO UPGRADE SRC-II

The refinery model combines the petroleum refinery with the SRC-II syncrude refinery by blending product streams to meet a given product slate. Additional process units are included where petroleum and SRC-II oil require separate treatment at different severity levels to meet product specifications. Based on a fixed feed of 50,000 BPD of SRC-II, and at a given production limit of gasoline, No. 2 fuel oil, No. 6 fuel oil, and turbine fuel, the program finds the most economical process route by reducing petroleum crude. To determine the impact of turbine fuel quality on the process economics, the process calculation for turbine fuel required selling price was determined at different nitrogen levels and endpoint specifications for turbine fuel.

A. Refinery Linear Programming Output Configuration

The refinery configurations, resulting from linear programming calculations, show two major processing modes. Case 1, Figure 6-16, shows the SRC-II oil being severely hydrotreated after vacuum distillation, then atmospherically distilled and blended with petroleum products. In Case 2, Figure 6-17, severe hydrotreating of fractions is applied, if necessary, after atmospheric distillation of SRC-II oil. For both cases, refinery process calculations were completed for the following turbine fuel specifications:

Case 1			
0.25% N		1.0% N	
TF1		--	distillate fuel
TF3	and	T13	heavy fuel

High nitrogen fuels are not achievable in Case 1 for distillate fuel because severe hydrotreating of whole SRC-II syncrude reduces the nitrogen content of distillate fractions below the turbine fuel specification.

Case 2			
0.25% N		1.0% N	
TF1		--	distillate fuel
TF2		--	distillate fuel with higher viscosity limit
TF3	and	T13	heavy fuel

Also in Case 2 high nitrogen fuels are not achievable for distillate turbine fuels because of the insufficient high nitrogen mid-distillates present in the syncrude refinery feed.

B. Calculation of Turbine Fuel Prices

To determine the required selling price of turbine fuel produced from a combined refinery consisting of petroleum and SRC-II syncrude feed, several basic operating conditions had to be defined, which are the same as applied to the previous synfuel cases and are as follows:

(1) The amount of gasoline was held constant since the market for this fuel does not change, disregarding normal seasonal variations.

(2) The amount of No. 2 fuel should stay constant but can be reduced.

(3) The amount of No. 6 fuel cannot be adjusted easily because of the heavier fuel oil content in the SRC-II syncrude. 24,965 BPD of No. 6 fuel oil are produced in Case 1 and 44,240 BPD are produced in Case 2, TF1, the difference being due to less hydrotreating in the latter case.

(4) 20,000 BPD of turbine fuel will be produced for the combined refinery cases.

(5) All product prices, except turbine fuel, remain the same. Thus turbine fuel price supports the profitability of the refinery expansion to meet a 15% discounted cash flow rate of return.

(6) The feed of SRC-II is fixed at 50,000 BPD, while crude oil feed can be reduced to meet the required product slate. The required product slate is defined with an upper limit for gasoline and No. 2 fuel oil, a lower limit for No. 6 fuel oil and a fixed turbine fuel production. There are no restrictions on the amount of other products.

The yield of No. 2 and No. 6 fuel oils in the combined refinery for petroleum plus SRC-II oil upgrading differs greatly from the parallel shale oil and H-Coal cases where distillate specifications for turbine fuels (TF1) are applied. This is mainly a result of the dissimilar feed of SRC-II oil which contains approximately 50% heavy resid in the boiling range over 950°F. This lack of sufficient middle distillate in the feed decreases the volume of potential No. 2 fuel blending stocks when 20,000 BPD of distillate turbine fuel production are required. Conversely, the No. 6 fuel oil blending stocks are proportionally increased which results in a high No. 6 fuel oil production when No. 2 fuel oil is at the specified minimal amount. This distorts the comparison with other synfuel refineries because of the different product values and the change in marketable products.

When the turbine fuel specifications are changed to those for heavier fuels, the amount of No. 2 fuel oil increases. However, the No. 6 fuel oil production decreases, but still shows a relative high No. 6 fuel oil production. The blending of SRC-II oil resid into No. 6 fuel oil and refinery fuel oil also increases the nitrogen content of these fuels to over 1 wt%.

The required turbine fuel selling prices results from different refinery configurations producing several grades of turbine fuel are shown in Table 6-26 and 6-27. Table 6-28 represents capital cost data for

the combined refinery with severe hydrotreating before atmospheric distillation, Case 1. Table 6-26 includes the calculated turbine fuel required selling prices for turbine fuels TF1, TF3 and T13 for Case 1.

Table 6-29 represents capital cost data for the combined refinery with hydrotreating after vacuum and atmospheric distillation, Case 2. Table 6-27 contains the calculated turbine fuel prices for turbine fuels TF1, TF2, TF3, and T13 for Case 2.

The "capital recovery factor" described in Section 6.3.1B for shale oil is used in calculating turbine fuel prices for SRC-II plus the existing petroleum refinery as shown in Tables 6-26 and 6-27.

The product slates for Cases 1 and 2 differ mainly in the amount of fuel oil produced for the several turbine fuel quality specifications and are shown in Tables 6-30 and 6-31.

C. Evaluation of Turbine Fuel Prices Versus Turbine Fuel Quality

The turbine fuel prices, Tables 6-26 and 6-27, reflect the effect of turbine fuel specification on the overall economics. The large amount of heavy resid in the SRC-II oil feed results in a high yield of No. 6 fuel oil which, because of the lower market value of No. 6 fuel oil, reduces the total refinery product value. Thus, the resid fraction has a major influence on the refinery economics when the turbine fuel specification is changed from distillate turbine fuel (TF1) to heavy turbine fuel (TF3) in both Cases 1 and 2.

In Case 1, where the SRC-II fraction C_4 to 950°F is hydrotreated and then further fractionated, a distillate and a heavy turbine fuel are produced, and also a heavy fuel with relaxed nitrogen specification. No distillate turbine fuel with nitrogen over 0.25 wt% was achievable. The change in turbine fuel specification from light to heavy fuel reflects in a price decrease of over 7%, while the change in nitrogen limit from 0.25 wt% to 1.0 wt% for the heavy fuel reduces the turbine fuel price approximately

5%. The other significant influence of specification change appears in the product amount increase of No. 2 and decrease of No. 6 fuel oils and the feed rate change of the petroleum crude. These fuel oil and feed quantity changes from TF1 to TF3 and T13 are shown in Table 6-30.

In Case 2, where no hydrotreating before fractionation takes place, distillate and heavy turbine fuels were produced and show a similar trend for increased endpoint and viscosity. The price for TF3 is more than 7% lower than for TF1, while the change in nitrogen limit from 0.25 wt% to 1% only shows a price decrease of 6% for TF3. Another influence appears to be the end point limitation for distillate fuels which shows a price change of approximately 6% when calculations are carried out with TF1 and TF2 specifications. Also in Case 2 the product amount of fuel oil and petroleum feed changes when different turbine fuel grades are produced. These changes are shown in Table 6-31.

The effect of change in nitrogen specification for TF3 is shown in Figure 6-18. A comparison of Case 1 and Case 2 turbine fuels with distillate and heavy fuel specifications and actual properties is shown in Figure 6-19.

D. Thermal Efficiency of Output Diagrams

The thermal efficiencies of the SRC-II oil plus existing petroleum refining for Cases 1 and 2 and each of the turbine fuel specifications are shown in Table 6-32. The thermal efficiencies for Case 1 turbine fuels TF1, TF3, and T13 are about 90.0% for all fuels. The thermal efficiencies for Case 2 turbine fuels TF1, TF2, and TF3, T13 are about 92.0%, or below, for all fuels cases.

E. Utilities of Output Diagram

The utilities requirements shown in Table 6-33 are based on providing 1,250 psig steam for driving letdown turbines to provide power requirements and low level process steam. Fuel is provided from refinery fuel

gas and fuel oil generated internally for firing heaters and boiler facilities. Cooling water, condensate, and sour water stripping facilities are also provided.

6.6.2 NEW SRC-II OIL REFINERY

Unlike the process calculation for the use of an existing refinery for SRC-II upgrading, the stand-alone SRC-II oil refinery is given full latitude in choosing a product slate, with the exception of 5,000 BPD of turbine fuel to be produced. LPG, gasoline, No. 2 and No. 6 fuel oil will be produced and blended to maximize the product value. With a fixed feed of 50,000 BPD SRC-II oil, the program finds the most economical process route, based on process yields and severity levels for the hydrotreatment of the different distillation fractions.

The refinery has to provide its own fuel for utility production. Hydrogen is produced from light gases from the refinery, but a unit for the partial oxidization of SRC-II resid to hydrogen is included to provide the hydrogen shortfall which cannot be produced from refinery streams.

To determine the impact of turbine fuel quality on the process economics, the linear program model was allowed to blend to different turbine fuel specifications, as described in the existing refinery cases.

A. Resulting Refinery Linear Programming Configuration

The refinery configurations represent economical process routes for upgrading SRC-II oil in a new refinery when different hydrotreating methods are applied. The difference between the two configurations, Figures 6-20 and 6-21, is the degree of hydrotreating before and after distillation of the 950°F minus fraction.

In Figure 6-20, Case 1, after separation of the 975°F plus resid, severe hydrotreating at high pressure and low space velocity occurs to hydrodenitrify the 975°F minus fraction to a nitrogen level of about 350 ppm

(wt). The result is an upgrading of 975°F minus fraction of SRC-II oil from an API of 18.6 to 30.0 degrees to a liquid suitable for further processing to petroleum specification products.

In Figure 6-21, Case 2, hydrotreating after distillation of the 950°F minus fraction takes place at high pressure and low space velocity to reduce the nitrogen to the level required to prevent poisoning and deactivation of the catalyst in subsequent processing units.

In none of the calculated Cases 1 and 2 was No. 2 fuel oil produced, due to the small amount of mid-distillate fraction available for turbine fuel, No. 6 fuel oil, and refinery fuel oil blending. In both of the calculated Cases 1 and 2, gasoline was produced as a means of increasing total product value. In both Cases 1 and 2, No. 6 fuel oil was blended from the 950°F plus fraction with hydrotreated gas oil and distillate streams to meet product specification No. 6 fuel oil for boiler feedstock.

In Figure 6-20, Case 1, four turbine fuels were produced: two distillate type turbine fuels, TF1 and TF2, with an endpoint of 650°F and less than 0.25 wt% nitrogen; and two wide range turbine fuels, TF3 and T13, with a greater than 1000°F endpoint with 0.25 wt% and 1.0 wt% nitrogen specification.

In Figure 6-21, Case 2, the linear program model was allowed to blend two different turbine fuel types: (1) TF1, a distillate turbine fuel with a 650°F endpoint, and (2) T12, a distillate turbine fuel like TF1 but allowing a higher viscosity, end point, and nitrogen content.

B. Calculation of Turbine Fuel Prices

To determine values for turbine fuels for the SRC-II oil refinery, the complete calculation was based on forcing the turbine fuel production of 5,000 BPD for Cases 1 and 2 at zero value. After deducting the daily capital recovery, operating cost and feed cost from the product value (excluding turbine fuel), a revenue margin was left which had to be supported

by the turbine fuel price. This price represents the value of turbine fuel to a SRC-II oil refinery forced to produce 5,000 BPD of turbine fuel. The turbine fuel price is based on selling all other products with petroleum specifications at the market prices prevailing for comparable petroleum products.

Table 6-34 presents capacity and capital cost data for the new SRC-II oil refinery for Case 1 which includes TF1, TF2, TF3 and T13 products, and Case 2 which includes TF1, TF2 and T12 products. Table 6-35 includes the required calculated turbine fuel prices for TF1, TF2, T12, TF3 and T13 for Cases 1 and 2 for the SRC-II oil refinery.

C. Evaluation of Turbine Fuel Prices Versus Turbine Fuel Quality

The evaluation of turbine fuel price calculations, as shown in Table 6-35, indicates the key factors that affect prices are as follows:

(1) The data in Table 6-35, severe hydrotreating before distillation, Case 1, TF1, TF2, TF3 and T13, indicate a high turbine fuel required selling price is required to support the 15% discounted cash flow profit level of the new SRC-II oil refinery. These prices are in a narrow range of about \$150-\$151 per barrel, or about 3.5 times the combined SRC-II plus petroleum refinery turbine fuel required selling price.

The factor that most affects this price difference is the high capital investment cost for the new SRC-II oil refinery. Unlike the existing refinery, where petroleum crude feed rate is reduced to allow existing process units to be used for SRC-II oil refining, all units must be sized and built specific to syncrude processing.

(2) The data pesented in Table 6-35 for Case 2, severe hydrotreating after distillation (TF1), indicates a lower capital investment cost for the new SRC-II oil refinery as compared with Case 1 but also a lower product value which results in a higher turbine fuel required selling price of about $155 per barrel. The major effect on the turbine fuel required selling price for TF1 is the change in product value and capital investment cost based on reducing the partial oxidation plant capacity due to decreased hydrogen consumption.

In Case 2, TF2, a much lower turbine fuel required selling price of $119 per barrel was calculated which results from deletion of the gas oil hydrotreater, hydrocracker, and coker units. This change in equipment requirement results from applying a higher endpoint specification for turbine fuel, TF2. Directionally, this specification change reflects the capital intensive changes that occur from deletion of refining units.

An evaluation of turbine fuel prices versus nitrogen level is shown in Figure 6-22 for Case 1 turbine fuels TF3 and T13. In Case 2, no heavy turbine fuels with TF3 specifications could be produced.

An evaluation of price versus endpoint specification for turbine fuels is shown in Figure 6-23 for Cases 1 and 2. These curves are a plot of the calculated turbine fuel prices versus a distillate type and a wide boiling range type of turbine fuel. The properties of the distillate and wide range turbine fuels are shown in the table below Figure 6-23. The results indicate that the fuel costs are insensitive to the product endpoint.

To produce the wide range turbine fuel, about 30% of the blended fuel has a boiling range over 650°F which results in a lower gravity and slightly higher viscosity of the product.

D. Thermal Efficiency

The thermal efficiencies of the SRC-II oil refinery for Cases 1 and 2 and each of the turbine fuel specifications are shown in Table 6-36. The thermal efficiencies for Case 1 turbine fuels TF1, TF2, TF3 and T13 averages about 86% for all fuels, while the thermal efficiencies for Case 2 turbine fuels TF1 and TF2 are about 87% and 96% respectively.

E. Utilities

The utilities requirement shown in Table 6-37 are based on providing a 1,250 psig steam plant for driving letdown turbines to provide power requirement and low level process steam. Fuel is provided from refinery fuel gas and fuel oil generated internally for firing heaters and boiler facilities. Cooling water, condensate, and sour water stripping facilities are also provided.

Table 6-26 - Turbine Fuel Selling Prices, Petroleum Crude Plus SRC-II Oil Refinery, 20,000 BPD Turbine Fuel Produced, $ per Day

Item	Case 1		
	TF1	TF3	TI3
Fixed Capital Investment for Additional Process Units	138,500,000	129,800,000	125,800,000
Add: Offsite Facilities $\frac{\text{FCI Add'tl Proc. Units} \times 0.30}{0.70}$	59,400,000	55,600,000	53,900,000
Royalties and Catalyst	2,997,000	3,543,000	3,116,000
Total Additional Capital Investment	200,897,000	188,943,000	182,816,000
Daily Capital Recovery (Total Add'tl Cap x 0.0010606*)	213,070	200,393	193,895
Add: Feed Cost	6,618,940	6,779,460	6,498,150
Operating Cost	57,222	57,519	56,362
Total Daily Required Revenue	6,889,232	7,037,372	6,748,407
Deduct: Product Values (Exclusive of Turbine Fuel)	6,684,560	6,898,870	6,655,410
Revenue Margin	204,678	138,502	92,997
Add: Operating Margin of Existing Refinery Before Addition of Synfuel Upgrading	702,204	702,204	702,204
Turbine Fuel Required Daily Revenue	906,882	840,706	795,201
Minimum Selling Price Per Barrel	45.34	42.04	39.76

* Capital Recovery Factor, stream day basis: $\frac{0.35}{\text{330 operating days per year}} = 0.0010606$

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Table 6-27 - Turbine Fuel Selling Prices, Petroleum Crude Plus SRC-II Oil Refinery, 20,000 BPD Turbine Fuel Produced, $ per Day

Item	Case 2 TF1	TF2	TF3	TI3
Fixed Capital Investment for Additional Process Units	55,200,000	57,900,000	83,700,000	74,800,000
Add: Offsite Facilities $\frac{\text{FCI Add'tl Proc. Units} \times 0.30}{0.70}$	23,660,000	24,800,000	35,800,000	32,060,000
Royalties and Catalyst	370,000	425,000	703,000	718,000
Total Additional Capital Investment	79,230,000	83,125,000	120,203,000	107,578,000
Daily Capital Recovery (Total Add'tl Cap x 0.0010606*)	84,031	88,162	127,487	114,097
Add: Feed Cost	7,155,700	6,905,600	6,842,200	6,802,160
Operating Cost	47,250	46,760	40,040	37,780
Total Daily Required Revenue	7,286,981	7,040,522	7,009,727	6,954,037
Deduct: Product Values (Exclusive of	7,146,400	6,949,100	6,929,600	6,923,200
Revenue Margin	140,581	91,422	80,127	30,837
Add: Operating Margin of Existing Refinery Before Addition of Synfuel Upgrading	702,204	702,204	702,204	702,204
Turbine Fuel Required Daily Revenue	842,785	793,626	782,331	733,041
Minimum Selling Price Per Barrel	42.14	39.68	39.11	36.65

* Capital Recovery Factor, stream day basis: $\frac{0.35}{\text{330 operating days per year}} = 0.0010606$

Table 6-28 - Capacity and Capital Cost Data, Petroleum Plus SRC-II Oil Refinery, Case 1

Process Units	Unit Capacities, BPD				Fixed Capital Investment ($ Million)		
	Existing Refinery	Equipment Additions					
		TF1	TF3	T13	TF1	TF3	T13
Crude Unit	200,000	—	—	—	—	—	—
Vacuum Distillation	75,000	—	—	—	—	—	—
Fluid Catalytic Cracker	50,000	—	—	—	—	—	—
Hydrocracker	10,300	—	—	—	—	—	—
Coker	12,500	4,830	—	—	16.4	—	—
Naphtha Hydrotreater	61,000	—	—	—	—	—	—
Atm Gas Oil Hydrotreater	22,000	—	—	—	—	—	—
Reformer	49,000	338	2,880	760	1.5	6.8	2.7
Alkylation	8,000	—		—	—	—	—
SRC-II Oil Vacuum Distillation	—	50,000	50,000	50,000	22.2	22.2	22.2
SRC Oil Atm Distillation	—	24,030	24,030	24,030	7.3	7.3	7.3
C_4-950°F Fraction Hydrotreater	—	25,500	25,500	25,500	76.7	76.7	76.7
Hydrogen Plant, million SCFD	—	14.73	13.54	15.88	8.8	8.3	9.2
Sulfur Recovery Plant, long ton/day	135	—	—	—	—	—	—
Ammonia Recovery from Waste Water, ton/day NH_3	17	—	—	—	—	—	—
Sour Water Stripper, M [illegible]/day	5,300	954	1,063	877	0.7	0.8	0.7
Cooling Water System, M gal/day	196,000	—	3,700	1,900	—	0.3	—
Steam/Power Plant, M lb/day, 1250 psig steam	15,100	177	462	335	2.1	4.6	4.2
Total Additional FCI					138.5	129.8	125.8

Table 6-29 - Capacity and Capital Cost Data, Petroleum Plus SRC-II Oil Refinery, Case 2

Process Units	Unit Capacity, BPD					Fixed Capital Investment ($ Million)			
	Existing Refinery	Equipment Additions							
		TF1	TF2	TF3	T13	TF1	TF2	TF3	T13
Crude Unit	200,000	—	—	—	—	—	—	—	—
Vacuum Distillation	75,000	—	—	—	—	—	—	—	—
Fluid Catalytic Cracker	50,000	—	—	—	—	—	—	—	—
Hydrocracker	10,300	—	—	—	1,490	—	—	—	14.1
Coker	12,500	—	—	—	200	—	—	—	2.0
Naphtha Hydrotreater	61,000	—	—	—	—	—	—	—	—
Atm Gas Oil Hydrotreater	22,000	—	—	—	—	—	—	—	—
Reformer	49,000	—	—	—	—	—	—	—	—
Alkylation	8,000	—	—	—	—	—	—	—	—
SRC-II Oil Vacuum Distillation	—	50,000	50,000	50,000	45,960	22.2	22.2	22.2	22.2
SRC-II Oil Atm Distillation	—	25,500	25,500	25,500	25,500	7.6	7.6	7.6	7.6
SRC-II Oil Naphtha Hydrotreater	—	5,840	5,840	5,840	5,840	24.8	24.8	24.8	24.8
SRC-II Oil Gas Oil Hydrotreater	—	—	666	4,301	—	—	2.5	25.0	—
Sulfur Recovery Plant, Long ton/day	135	—	—	—	2	—	—	—	0.3
Ammonia Recovery from Waste Water, ton/day NH_3	17	1	2	9	1	0.3	0.5	0.3	0.3
Sour Water Stripper, M lb/day	5,300	368	343	444	443	0.3	0.3	0.4	0.4
Cooling Water System, M gal/day	196,000	—	—	—	—	—	—	—	—
Steam/Power Plant, M lb/day, 1250 psig steam	15,100	—	—	324	287	—	—	3.4	3.1
Total Additional FCI						55.2	57.9	83.7	74.8

Table 6-30 - Product Slate, SRC-II Oil Plus Petroleum Refinery, Case 1

Item	Rate	TF1	TF3	T13
Feed				
Petroleum Crude	M BPD	170.63	175.98	166.6
SRC-II Oil	M BPD	50.0	50.0	50.0
Products				
LPG	M BPD	11.09	12.55	8.42
Gasoline	"	108.1	108.1	108.1
No. 2 fuel oil	"	45.8	53.8	53.8
No. 6 fuel oil	"	24.965	21.739	15.653
Turbine fuel	"	20.0	20.0	20.0
Sulfur	M LTPD	0.12	0.13	0.124
Ammonia	M TPD	0.054	0.054	0.053
Coke	M TPD	1.009	0.749	0.778

Table 6-31 - Product Slate, SRC-II Oil Plus Petroleum Refinery, Case 2

		Turbine Fuels			
Item	Rate	TF1	TF2	TF3	T13
Feed					
Petroleum Crude	M BPD	188.52	188.2	178.07	176.74
SRC-II Oil	M BPD	50.0	50.0	50.00	50.0
Products					
LPG	M BPD	12.332	9.296	9.032	9.357
Gasoline	"	108.1	108.1	108.1	108.1
No. 2 fuel oil	"	48.8	47.264	52.7	53.8
No. 6 fuel oil	"	44.24	36.903	28.569	26.485
Turbine fuel	"	20.0	20.0	20.0	20.0
Sulfur	M LTPD	0.128	0.128	0.139	0.137
Ammonia	M TPD	0.018	0.019	0.027	0.018
Coke	M TPD	0.683	0.689	0.772	0.804

Table 6-32 - Thermal Efficiencies of SRC-II Plus Existing Petroleum Refinery

Item	Million Btu per Day						
	Case 1			Case 2			
	TF1	TF2	TF3	TF1	TF2	TF3	T13
Total Heating Value Feed	1306.2	1337.1	1282.9	1409.5	1344.7	1342.5	1341.5
Total Heating Value Products	1181.3	1208.3	1154.6	1298.3	1230.7	1228.9	1227.9
Thermal Efficiency, %	90.4	90.4	90.0	92.1	91.5	91.5	91.5

Table 6-33 - Total Utilities Requirement, SRC-II Plus Existing Petroleum Refinery (Computer Output)

Units	Usage Rate	
	Case 1	Case 2
Sour water stripping	6254 M lb/D	5668 M lb/D
Cooling water circulation	191 MM gal/D	184 MM gal/D
Power generation	1255 M kWh/D	1205 M kWh/D
Fuel consumption	81.5 MMM Btu/D	78.7 MMM Btu/D

Table 6-34 - Capacity and Capital Cost Data, New SRC-II Oil Refineries, Cases 1 and 2

Process Units	Unit Capacity, BPD							Fixed Capital Investment, $ Million						
	Case 1				Case 2			Case 1				Case 2		
	0.25%N TF1	0.25%N TF2	0.25%N TF3	1.00%N T13	0.25%N TF1	0.25%N TF2	1.00%N T12	TF1	TF2	TF3	T13	TF1	TF2	T12
Vacuum Splitter	50,000	50,000	50,000	50,000	50,000	50,000	50,000	22.2	22.2	22.2	22.2	22.2	22.2	22.2
Atm Distillation	24,030	24,030	24,030	24,030	25,500	25,500	25,500	7.3	7.3	7.3	7.3	7.6	7.6	7.6
950°F- Hydrotreater	25,500	25,500	25,500	25,500	—	—	—	76.7	76.7	76.7	76.7	—	—	—
Naphtha Hydrotreater	8,650	8,650	8,650	8,650	6,126	5,840	5,840	31.4	31.4	31.4	31.4	25.5	24.8	24.8
Gas Oil Hydrotreater	—	—	—	—	9,245	3,310	—	—	—	—	—	41.7	—	—
Hydrocracker	3,740	5,040	6,500	6,900	6,490	—	—	24.5	29.3	34.2	35.4	34.1	—	—
Coker	—	—	—	—	2,649	—	—	—	—	—	—	11.4	—	—
Reformer	10,190	10,730	11,310	10,200	8,059	4,964	3,888	16.6	17.2	17.8	16.6	14.1	10.0	8.4
Partial Oxidation Plant, MM SCFD	35.0	35.3	35.6	35.7	38.2	7.2	—	32.6	32.9	33.1	33.2	35.5	8.0	—
Sulfur Recovery Plant, long ton/day	12	12	12	12	24	17	15	1.0	1.0	1.0	1.0	1.6	1.3	1.1
Ammonia Recovery from Waste Water, ton/day NH_3	38	38	38	38	30	8	1	2.8	2.8	2.8	2.8	2.4	1.0	0.3
Sour Water Stripper, M lb/day	4,092	4,202	4,323	4,359	4,036	1,274	704	2.4	2.4	2.5	2.5	2.3	1.2	0.6
Cooling Water System, M gal/day	40,780	42,600	44,600	45,200	42,717	13,600	7,100	1.7	1.8	1.8	1.9	1.8	0.8	0.4
Steam/Power Plant, M lb/day, 1250 psig steam	7,643	7,931	8,249	8,343	7,895	2,272	1,160	43.5	44.8	46.2	46.6	44.6	15.0	9.6
Total Fixed Capital Investment								262.7	269.8	277.0	277.6	244.8	111.9	75.0

Table 6-35 - Turbine Fuel Selling Prices, New SRC-II Oil Refinery, 5,000 BPD Turbine Fuel Produced, $ Per Day

Item	Case 1				Case 2		
	TF1 (0.25% N) 650°F EP	TF2 (0.25% N) Below 1000°F	TF3 (0.25% N) Above 1000°F	T13 (1.00% N) Above 1000°F	TF1 (0.25% N) 650°F EP	TF2 (0.25% N) Below 1000°F	T12 (1.00% N) Below 1000°F
Fixed Capital Investment for Process Units	262,700,000	269,800,000	277,000,000	277,600,000	244,800,000	111,900,000	75,000,000
Add: Offsite Facilities FCI Add'tl Proc. Units x 0.30 / 0.70	112,586,000	115,629,000	118,714,000	118,971,000	104,900,000	47,760,000	32,100,000
Royalties and Catalyst	7,750,000	8,170,000	8,620,000	8,720,000	7,590,000	2,190,000	1,199,000
Total Capital Investment	383,036,000	393,599,000	404,334,000	405,291,000	357,290,000	162,050,000	108,299,000
Daily Capital Amortization (Total Add'tl Cap x 0.0010606)	406,248	417,451	428,837	429,852	378,942	171,870	114,862
Add: Feed Cost	1,500,000	1,500,000	1,500,000	1,500,000	1,500,000	1,500,000	1,500,000
Operating Cost	4,970	5,170	5,380	5,450	25,005	14,730	14,730
Total Daily Cost	1,911,218	1,922,621	1,934,217	1,935,302	1,903,947	1,686,600	1,629,592
Deduct: Product Values (Exclusive of Turbine Fuel)	1,156,000	1,168,000	1,182,000	1,185,000	1,129,500	1,093,300	1,093,300
Turbine Fuel Required Daily Revenue	755,218	754,621	752,217	750,302	774,447	593,300	536,292
Minimum Selling Price Per Barrel Turbine Fuel	151.04	150.92	150.44	150.06	154.89	118.66	107.26

Table 6-36 - Thermal Efficiencies of New SRC-II Oil Refinery

Item	Million Btu per Day						
	Case 1				Case 2		
	TF1	TF2	TF3	T13	TF1	TF2	T12
Total Heating Value Feed	320.6	320.6	320.6	320.6	320.6	320.6	320.6
Total Heating Value Products	277.4	276.5	275.5	275.2	278.2	300.7	307.9
Thermal Efficiency, %	86.5	86.2	85.9	85.8	86.8	93.8	96.0

Table 6-37 - Total Utilities Requirement, New SRC-II Oil Refinery

Unit	Usage Rate	
	Case 1	Case 2
Sour Water Stripping	4092 M lb/D	4036 M lb/D
Cooling Water Circulation	41 MM gal/D	43 MM gal/D
Power Generation	459 M kWh/D	488 M kWh/D
Fuel Consumption	18.1 MMM Btu/D	18.5 MMM Btu/D

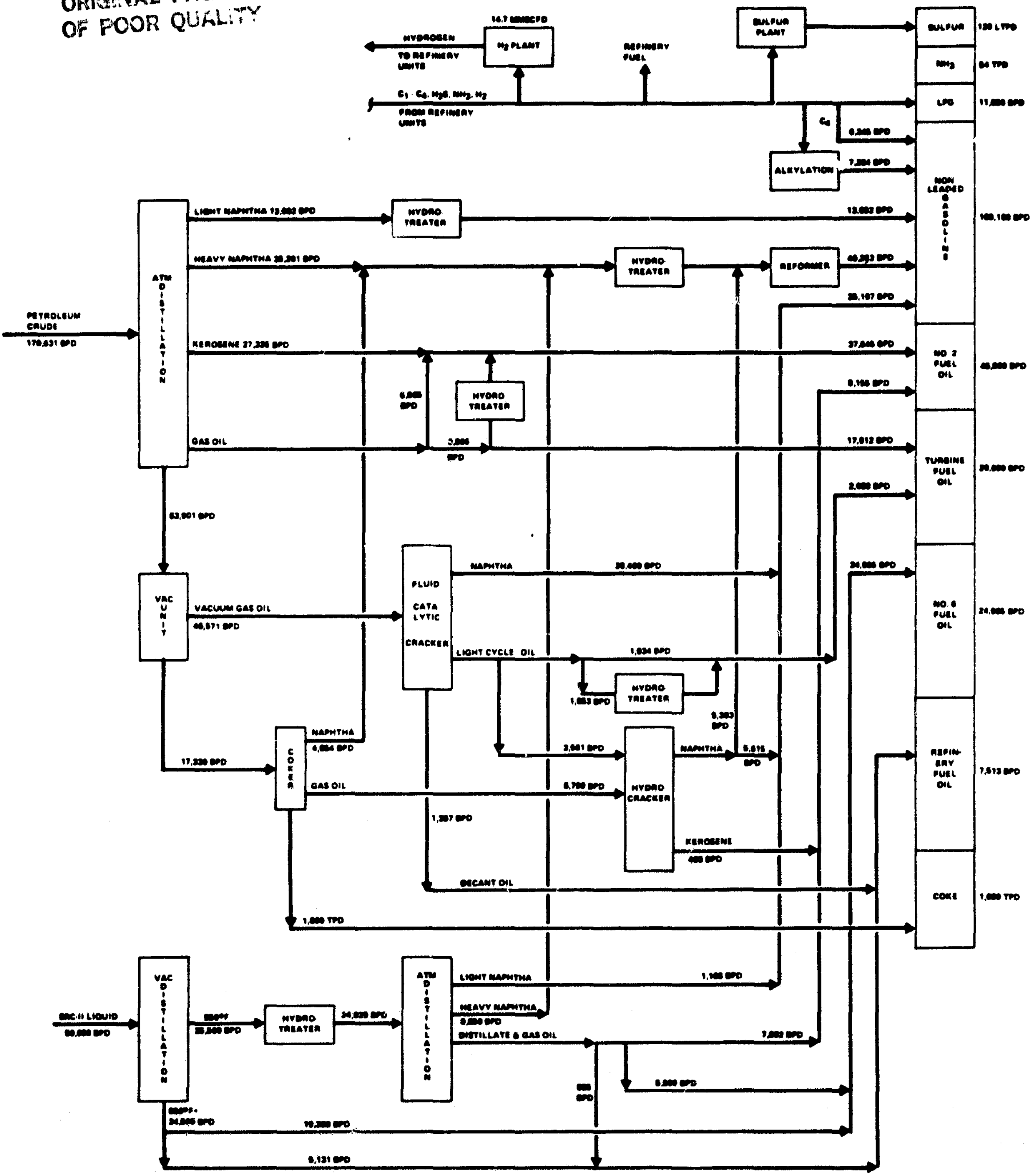


Figure 6-16 - Computer Output Data Diagram, Petroleum Refinery Plus SRC II Refinery, Case 1

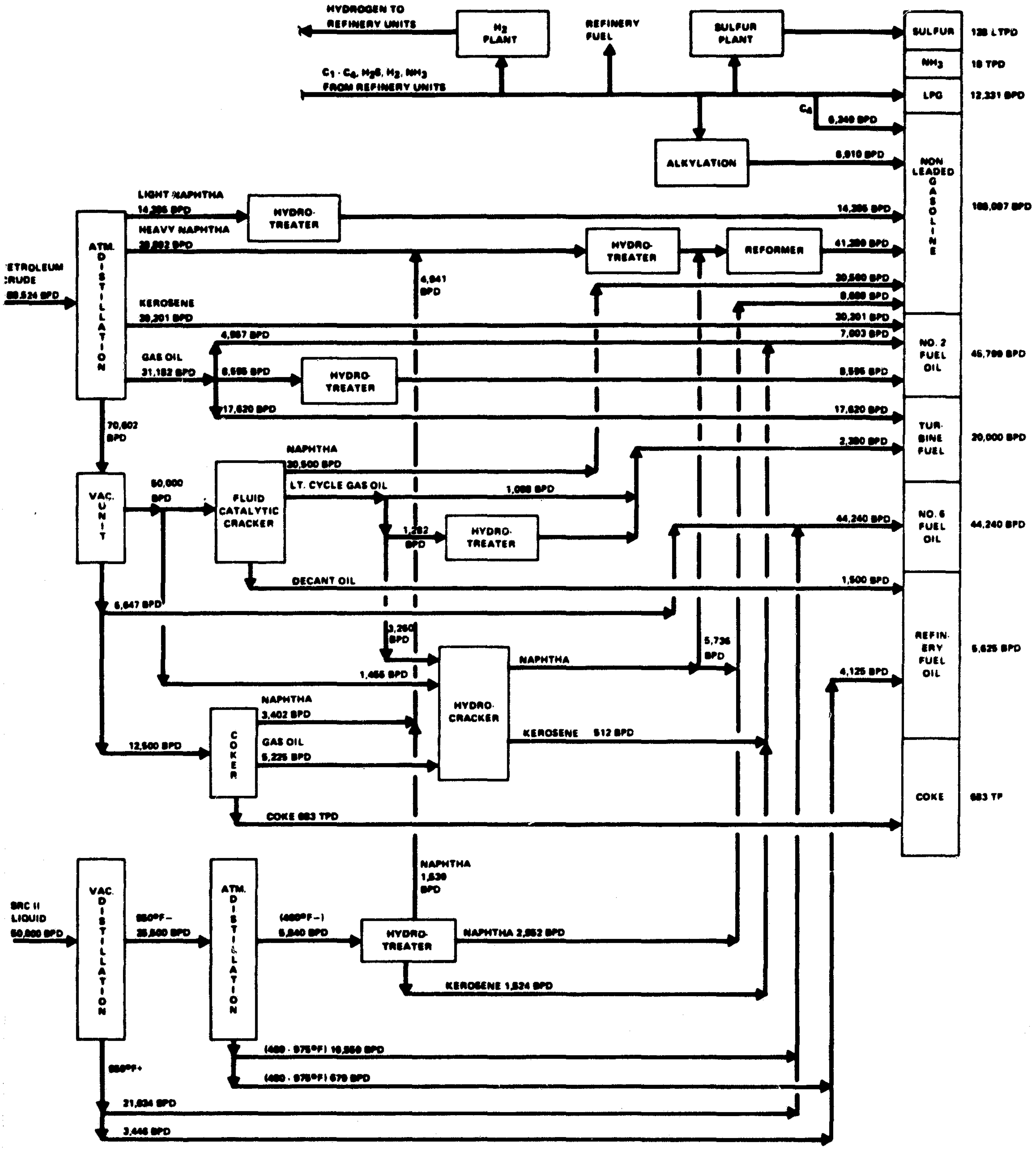
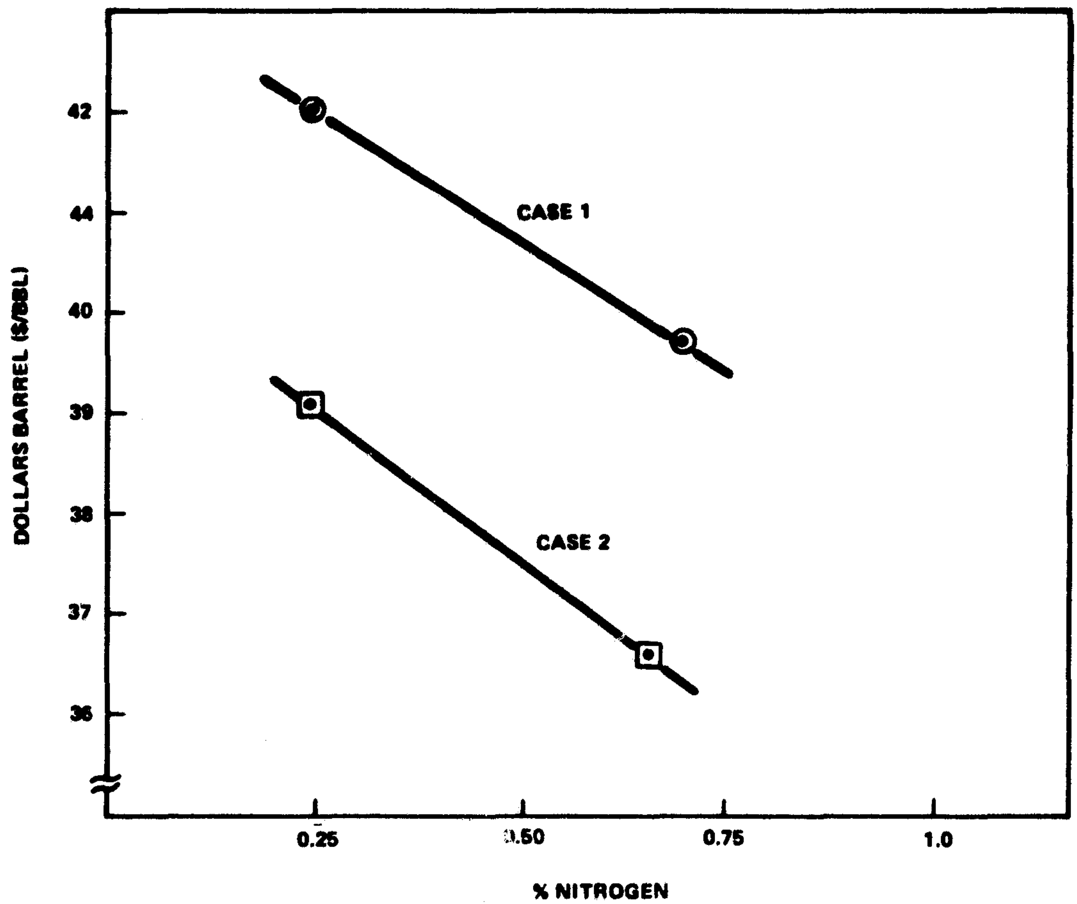


Figure 6-17 - Computer Output Data Diagram, Petroleum Refinery Plus SRC II Refinery, Case 2

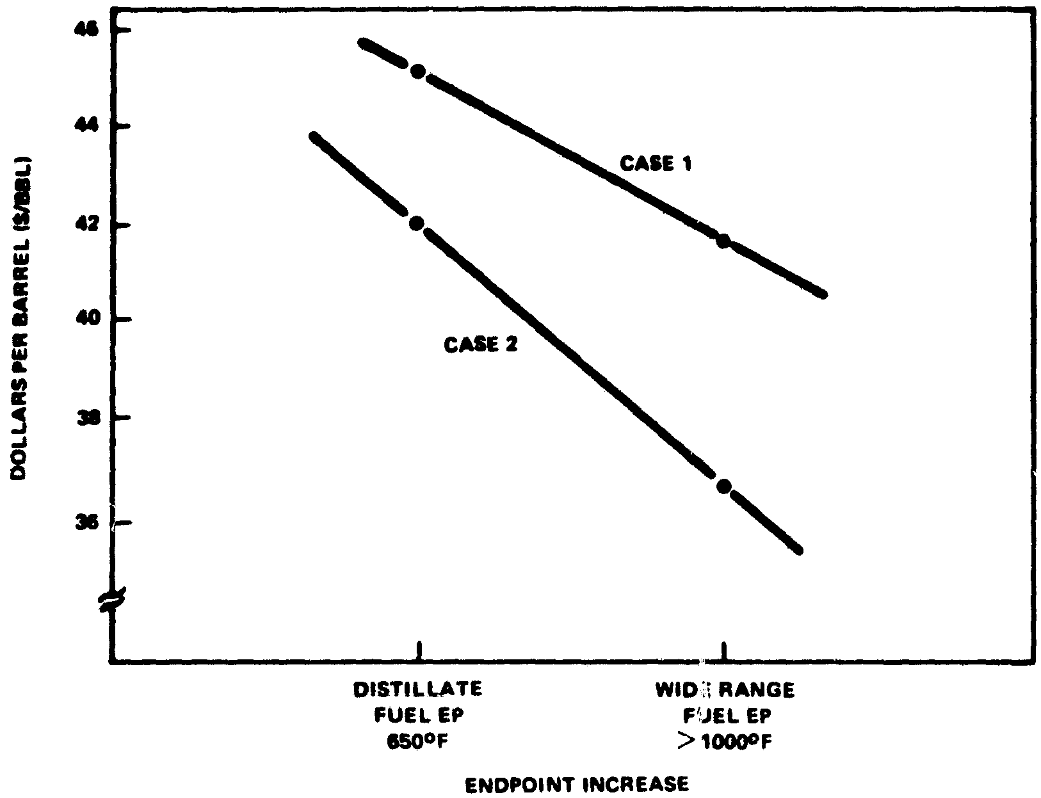


LEGEND:

◎ TURBINE FUEL 3 (SPECIFICATION TF3, TABLE 6-1), FIGURE 6-16

⊡ TURBINE FUEL 3 (SPECIFICATION TF3, TABLE 6-1), FIGURE 6-17

Figure 6-18 - Effect of Varying the Nitrogen Specification of Turbine Fuel on Price, SRC II Liquid Plus Existing Petroleum Refinery, Cases 1 and 2



PROPERTIES OF TURBINE FUELS:

PROPERTY		TYPE OF FUEL							
		DISTILLATE CASE 1		WIDE RANGE FUEL CASE 1		DISTILLATE CASE 2		WIDE RANGE FUEL CASE 2	
		ACTUAL	SPECIFICATION TF1[a]	ACTUAL	SPECIFICATION TF3[a]	ACTUAL	SPECIFICATION TF1[a]	ACTUAL	SPECIFICATION TF3[a]
GRAVITY, °API	(MIN)	32.300	15.00	16.70	15.00	31.900	15.00	17.60	15.00
SULFUR, WT%	(MAX)	.700	0.70	0.70	0.70	0.700	0.70	0.70	0.70
NITROGEN, WT%	(MAX)	.036	0.25	0.25	1.25	0.038	0.25	0.25	0.25
VISCOSITY (100°F), cst	(MAX)	5.400	5.80	21.00	[illegible]	5.200	5.80	12.00	160.00
FRACTION BOILING OVER 650°F, %		0.000	0.00	40.00	≤100.00	0.000	0.00	40.00	≤100.00

[a]SEE TABLE 6-1.

Figure 6-19 - Effect of Varying the Endpoint Specification of Turbine Fuel on Price, SRC II Liquid Plus Existing Petroleum Refinery, Cases 1 and 2

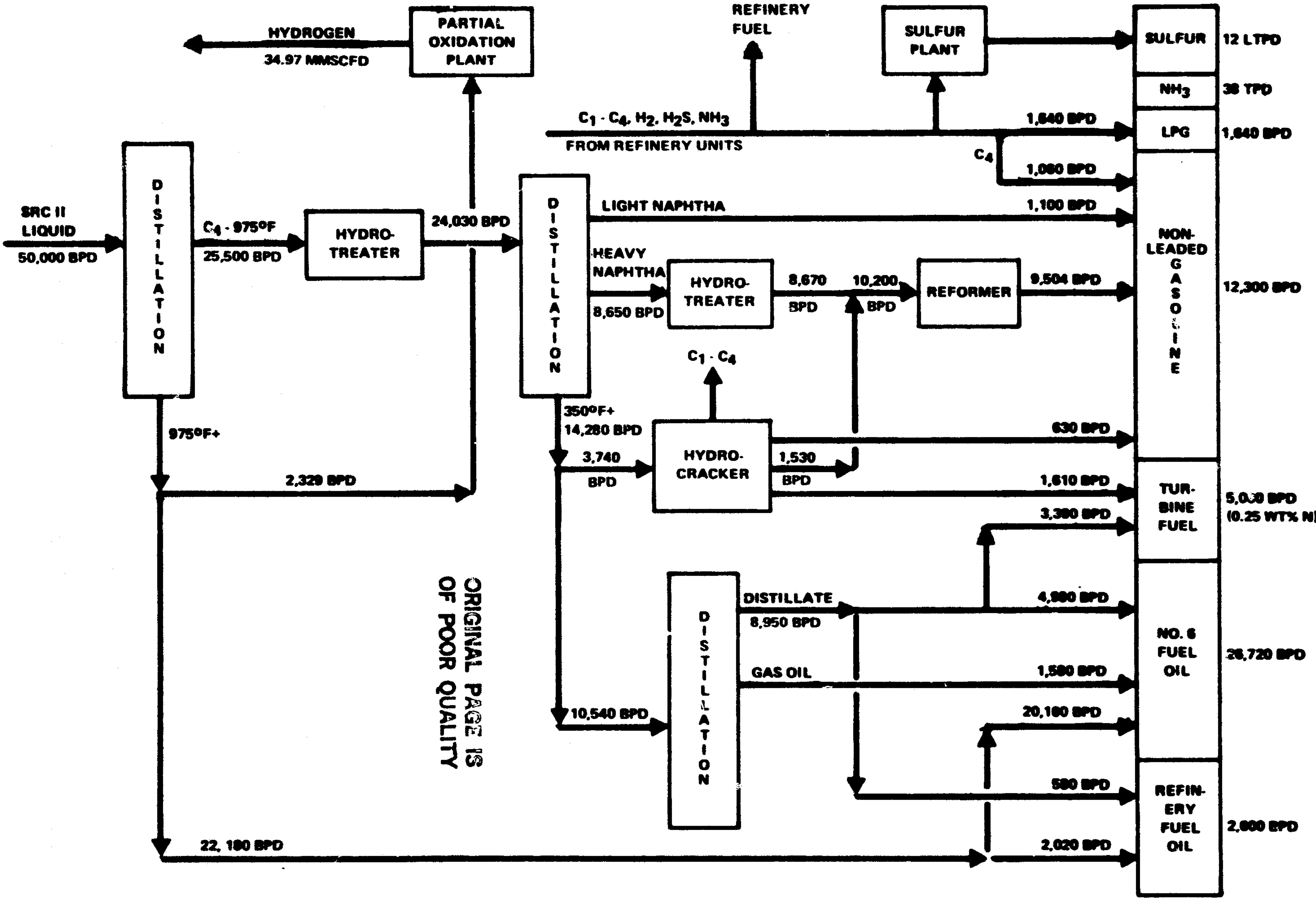


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Figure 6-20 - Computer Output Data Diagram, New SRC II Refinery, Case 1

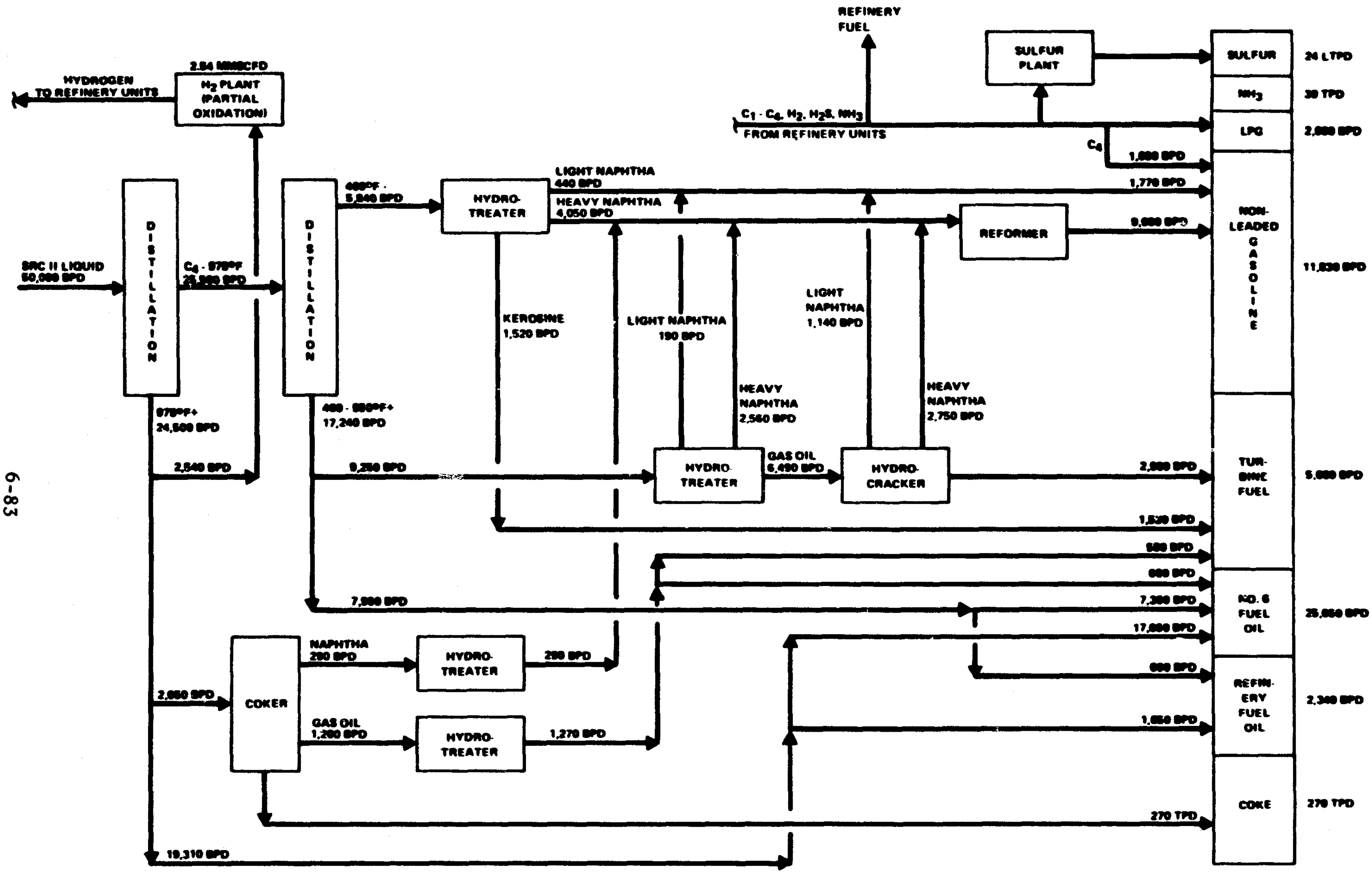


Figure 6-21 - Computer Output Data Diagram, New SRC II Refinery, Case 2

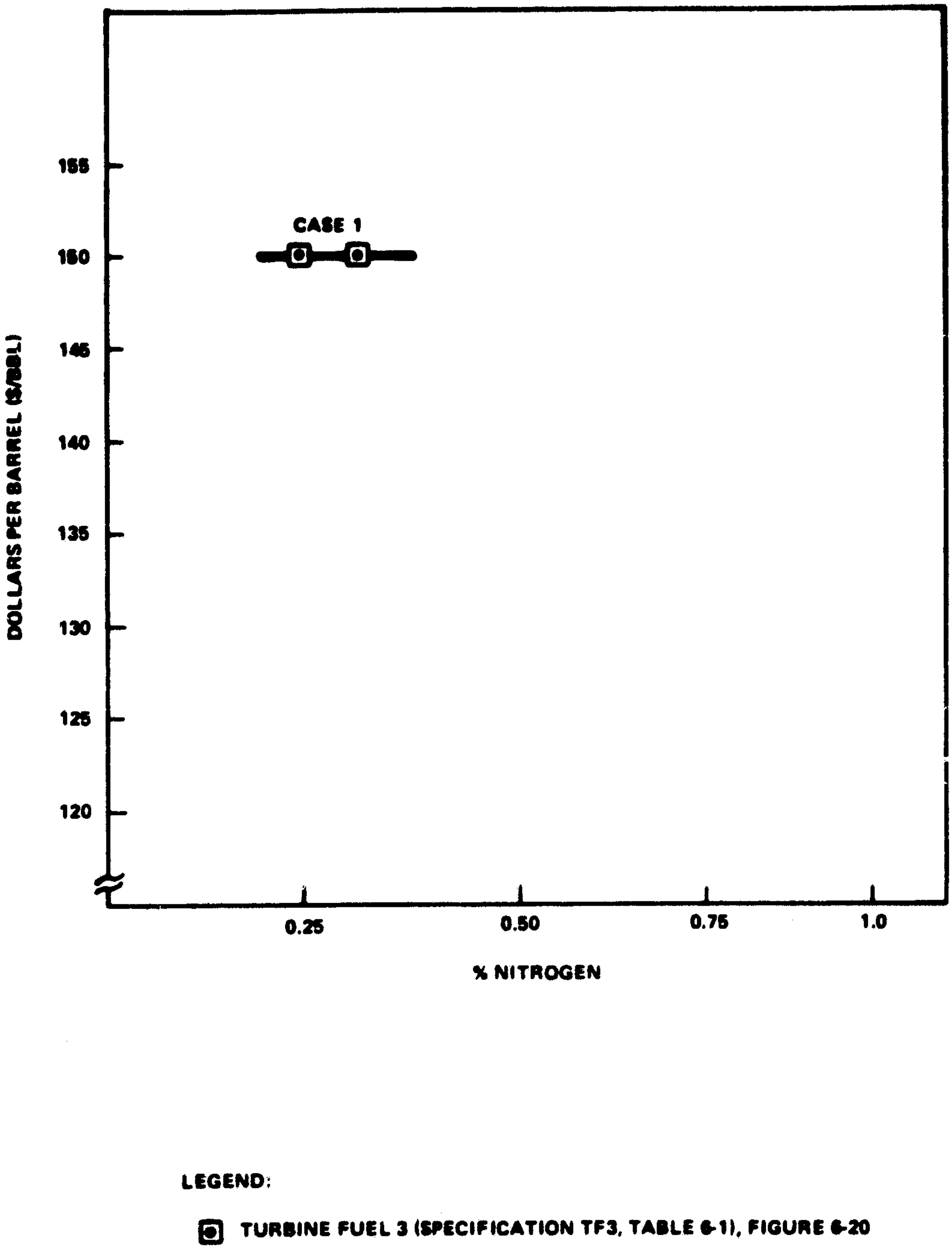
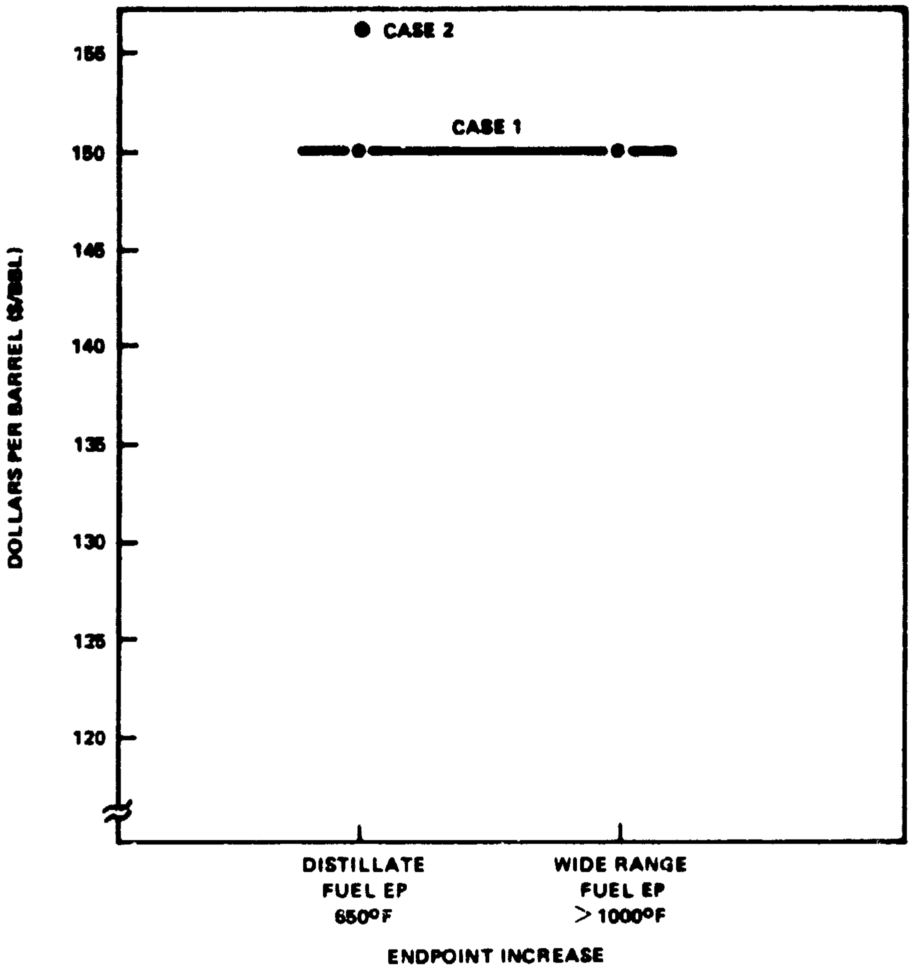


Figure 6-22 - Effect of Varying the Nitrogen Specification of Turbine Fuel on Price, New SRC II Refinery, Case 1



PROPERTIES OF TURBINE FUELS:

PROPERTY		DISTILLATE CASE 1		WIDE RANGE FUEL CASE 1		DISTILLATE CASE 2	
		ACTUAL	SPECIFICATION TF1*	ACTUAL	SPECIFICATION TF3*	ACTUAL	SPECIFICATION TF1*
GRAVITY, °API	(MIN)	23.400	15.00	16.20	15.00	26.900	15.00
SULFUR, WT%	(MAX)	0.002	0.70	.07	0.70	.005	0.70
NITROGEN, WT%	(MAX)	0.070	0.25	0.25	0.25	.028	0.25
VISCOSITY (100°F), cst	(MAX)	2.400	5.80	10.00	160.00	3.500	5.80
FRACTION BOILING OVER 650°F, %		0.000	0.00	30.00	≤ 100.00	0.000	0.00

*SEE TABLE 6-1.

Figure 6-23 - Effect of Varying the Endpoint Specification of Turbine Fuel on Price, New SRC-II Refinery, Cases 1 and 2

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6.7 CONCLUSIONS

A summary of conclusions regarding the material included in the previous sections is presented.

6.7.1 UPGRADING SYNCRUDES IN EXISTING REFINERIES

The evaluation of the synfuel upgrading by feeding an existing refinery indicates a substantial reduction of petroleum crude feed is possible. The reduction of 30,000 to 40,000 BPD from the charging of 50,000 BPD of syncrudes had been calculated. This may be thought of as an equivalent reduction in foreign crude oil imports.

6.7.2 COMPARATIVE FCI AND PRODUCTION COSTS

The results of this assessment indicate that the processing of synthetic crudes in an existing refinery with petroleum at a reduced feed rate is the most economical route. Fixed capital investments as well as production costs for processing synfuels are lower than those for a new smaller syncrude refinery. The following relationships summarize this situation for the three synfuel feeds studied. The results are expressed as averages of the cases studied. Each case processed 50,000 BPD of syncrude.

	Feed Cost ($/bbl)	FCI Average ($ Million) Existing[a] Refinery	FCI Average ($ Million) New[b] Refinery	Turbine Fuel Required Revenue ($ per barrel) Existing Refinery	Turbine Fuel Required Revenue ($ per barrel) New Refinery
Shale Oil	25	215	488	32	103
H-Coal Oil	32	121	247	40	101
SRC-II Oil	30	95	213	41	119
Petroleum Crude	30	--	--	--	--

[a] FCI of process unit additions to a 200,000 BPD petroleum refinery having a base FCI of approximately $600 million.

[b] FCI of process units for refining 50,000 BPD of syncrudes.

Although first inspection indicates that FCI is the dominant factor, sensitivities summarized in Section 6.7.4 indicate that the feed cost is a significant factor in determination of the eventual production costs which determines the revenue requirement. Certain modes of operation are more advantageous than others. This factor is discussed in the following subsection

6.7.3 UPGRADING REFINERY CASE DEFINITIONS

Two major approaches are described for upgrading of synfuels in combination with a petroleum refinery and in a stand-alone refinery.

In Case 1, for shale oil, H-Coal and SRC-II liquids, severe hydrotreating of the whole liquid (C_4 to approximately 950°F for SRC-II) was assessed. In Case 2, for the same liquids, severe hydrotreating of the individually distilled fractions was applied. Using Cases 1 and 2, a wide range of synfuel upgrading possibilities could be considered for fixed and undetermined product slates and quantities.

The turbine fuels produced by the different refinery complexes and different crude feeds are mainly classified as distillate fuels and heavy/residual fuels. These are distinguished by endpoint restrictions for distillates, and no endpoint limit and high viscosity for heavy fuels. These fuels were produced with a maximum nitrogen content of 0.25 wt% and a maximum nitrogen content of 1.0 wt%.

These turbine fuels are within the specifications defined in Table 6-1. Existing limited data hampered assessment of the effect of refining on metals content. Additional metals content data is necessary. Since the turbine fuels have gone through several catalytic processing steps before being blended to turbine fuel, a low metals content would be expected due to deposition in the catalyst beds. Only in the case of SRC-II, where a heavy turbine fuel is produced, is the heavy resid from the liquefaction process blended into the turbine fuel. This fraction may contain a large amount of impurities which could be unacceptable as turbine fuel. Production

and analysis of this proposed turbine fuel will be required to determine its acceptability.

In most of the cases where turbine fuel is a product of the combined petroleum/synfuel refinery, the blending stocks for turbine fuel originate mainly from the petroleum gas oil. The synfuel fractions in the combined refinery are mainly blended into fuel oils other than turbine fuels. This is a result of the internal value of different stocks in the refinery.

In the stand-alone syncrude refinery, the nitrogen limit results in upgrading of the fractions and thereby reduces the metal impurity level as well due to catalyst contact. The only exception is the relatively untreated SRC-II liquid which in some cases is blended directly into the turbine fuel pool. Quantification will be possible through future experience.

In a comparison of different methods for refining syncrudes in existing refineries and stand-alone synfuel refining to produce marketable products and additional turbine fuel, the turbine fuel value reflects the relative economics of the separate process configurations.

A. Use of Existing Refinery to Upgrade Synfuels

Of all the process calculations and economics evaluations, Case 2, with minimum or no hydrotreating of the whole syncrude, shows the lowest turbine fuel prices. This indicates that the hydrotreating of whole syncrude, as in Case 1, is more severe than product specifications demand. Further, an increase in turbine fuel endpoint and nitrogen content reduces the fuel cost significantly. Accordingly, hydrotreating should be held to the minimum level necessary to achieve required specifications. However, hydrotreating is necessary to reduce the nitrogen content to the point where catalyst poisoning does not occur in subsequent hydrocracking and FCC operations.

The range of turbine fuel prices for these combined process configurations is between \$29 and \$45 per barrel. The turbine fuels derived

from shale oil, based on the feedstock values and other parameters used in this study, are in the lower part of this range and show less upgrading costs. Turbine fuels from H-Coal and SRC-II are in the range of $36 to $45 per barrel reflecting the more severe upgrading necessary for coal liquids to compete with petroleum liquids.

These prices are relative figures which depend strongly on the feedcost and product values used in the economic evaluation.

B. Synfuel Refinery

The turbine fuel prices from stand-alone synfuel refineries show a trend similar to the combined refinery. The range of turbine fuel price is from $98 to $155 per barrel, with the exception of H-Coal in Case 2, which produced over 6,400 BPD turbine fuel and so reduced the turbine fuel price.

In nearly all the other calculations, hydrotreating after distillation (Case 2) shows a lower cost figure than when hydrotreating the syncrude before distillation (Case 1). Shale oil products were in a lower cost range, between $98 and $116 per barrel, and SRC-II products were close to $150 per barrel with a strong decrease in price for Case 2 when the specifications were relaxed.

6.7.4 SENSITIVITY TO CAPITAL INVESTMENT AND SYNCRUDE COST

The sensitivity of required turbine fuel selling prices to plus and minus 30% changes in total capital investment and to syncrude feed costs were developed for each of the three syncrude feeds for Case 2, turbine fuel TF1. Comparisons of the base value of the estimated required product selling prices (RPSP), presented in Section 6 with comparable values when the investment and feed costs are independently varied $\pm$30%, are presented in Table 6-38. The tabulations indicate that a given percentage change in syncrude feed price has a greater effect on the required product selling prices than a similar percentage change in the total capital investment.

Tables 6-39 and 6-40 summarize sensitivity ratios of RPSP to capital investment and synfuel feed cost. Again, these results show the high sensitivity to feedstock cost. For example, a 10 percent change in SRC-II feedstock cost results in a $30 per barrel change in TF1 turbine fuel cost.

Availability of sensitivity values as presented in Tables 6-38, 6-39 and 6-40 will permit the reader flexibility in interpreting the results presented in this report. To expedite our analysis, specific synfuel feedstock costs were selected based on publicly available estimates. In a sense, these selected feedstock costs represented judgment but nevertheless somewhat arbitrary decisions. The availability of the sensitivity values will permit the reader to quickly and independently select an alternative feedstock cost and determine the impact of this alternative value on the refinery economics. Similarly, the effects of variations in refinery fixed capital investments can be quickly estimated.

Table 6-38 - Required Product Selling Price Sensitivities
Case 2, Turbine Fuel TF1
$ per barrel

Syncrude Feed	Existing Refinery -30%	Existing Refinery Base Case	Existing Refinery +30%	New Syncrude Refinery -30%	New Syncrude Refinery Base Case	New Syncrude Refinery +30%
Sensitivity to Total Capital Investment:						
Shale Oil	29	34	40	57	103	149
H-Coal	37	39	41	49	67	84
SRC-II	41	42	43	132	155	178
Sensitivity to Syncrude Feed Cost:						
Shale Oil	15	34	54	26	103	180
H-Coal	15	39	63	- 8	67	141
SRC-II	20	42	65	65	155	245

Table 6-39 - Sensitivity Ratios of Turbine Fuel Required Product Selling Price (RPSP) to Fixed Capital Investment (FCI)

Syncrude	Existing Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% FCI}}$	New Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% FCI}}$
Shale Oil	0.183	1.53
H-Coal	0.067	0.58
SRC-II	0.033	0.77

Table 2-5 - Sensitivity Ratios of Turbine Fuel Required Product Selling Price (RPSP) to Syncrude Feed Cost

Syncrude	Existing Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% Syncrude Cost}}$	New Refinery - Sensitivity in $\frac{\Delta \text{ RPSP (\$/bbl)}}{\Delta \text{ \% Syncrude Cost}}$
Shale Oil	0.65	2.57
H-Coal	0.80	2.48
SRC-II	0.75	3.00

SECTION 7

ENVIRONMENTAL

The pollutants emitted from gas turbines are those common to all combustion sources: particulates, hydrocarbons (HC), carbon monoxide (CO), sulfur dioxide (SO_2), and nitrogen oxides (NO_x). The mass emissions from stationary gas turbines will differ depending on several variables such as turbine firing temperature, turbine pressure ratio, turbine load, combustor design, and atmospheric conditions.[1]

7.1 STANDARDS

The U.S. Environmental Protection Agency has issued New Source Performance Standards[2] for stationary gas turbines as follows:

Sulfur Dioxide: maximum emissions of 150 ppm or use of fuel containing a maximum sulfur content of 0.8% by weight.

Nitrogen Oxides (as nitrogen dioxide):

Gas turbines of heat input greater than 100 MM Btu/hr:	75 ppm
Gas turbines of heat input included between 10 and 100 MM Btu/hr:	150 ppm

Additional allowance for fuel bound nitrogen: up to 50 ppm for nitrogen content of 0.25% or higher.

Additional allowances are provided for thermal efficiencies greater than 25%; emissions are based on 15% oxygen content, no water present.

It is assumed that large turbines can meet the 75 ppm limit by injection of water or steam, while smaller turbines can meet the 150 ppm limit using dry controls. The fuel bound nitrogen allowance of additional 50 ppm of NO_2 may permit use of fuels containing a maximum 0.25% nitrogen content.[1,2,3]

While EPA has not issued emission standards for the other pollutants, emissions have to meet ambient air quality standards after dilution from atmospheric dispersion.

Occupational Safety and Health (OSHA) standards also must be met within the battery limits of gas turbine plants. Of particular interest when firing synfuels is the OSHA Standard of 0.2 mg/m^3 (8-hour average) for coal tar pitch volatiles (anthracene, benzo(a)pyrene, phenanthrene, acridine, chrysene, and pyrene).

7.2 PARTICULATE EMISSIONS

Particulate emissions are defined as "solid or liquid particles suspended in air with the exception of water in all its physical forms." Particulate emissions from gas turbines consist of ash from the fuel, carbon particles and hydrocarbons resulting from incomplete combustion. Fuels containing high ash and vanadium contents, such as crude or residual fuels, will result in higher particulate emission rates than light distillate fuels or natural gas. Particulate emissions may be decreased by combustor modifications which provide more complete combustion of hydrocarbons and carbonaceous particles.

Specific aspects of particulate emissions are their increased hazardous nature when consisting of high boiling hydrocarbons (see Section 7.6), and the persistent visibility ("smoke") of the small-size fraction (particle diameter of less than one micron). The latter effect is due to increased light scattering by particles with diameters of the same order of magnitude as the wave length of visible light.

Paraffinic saturated fuels tend to "smoke" less than the aromatic or unsaturated fuels and this smoking tendency is related to the chemical bond energies necessary to completely consume the fuel. Fuel hydrogen content and residual carbon content also affect visible emissions. A reduction in hydrogen content or an increase in residual carbon, or both, can increase visible emissions. Major reductions in visible emissions have been achieved

through combustor redesign to provide more effective fuel and air mixing in the primary zone and sufficiently lean regions within the combustor for smoke burnout.

7.3 HYDROCARBONS AND CARBON MONOXIDE EMISSIONS

Incomplete combustion is the principal cause of emissions of hydrocarbons (HC) and carbon monoxide (CO). Gas turbines are typically designed for optimum combustion efficiency in excess of 99% at full load. This efficiency, however, may drop to the 90 to 95 percent range for operation at idle or low power conditions. Because of this drop, emissions of HC and CO from the turbines will be higher for turbine start-up and operation at low loads and will be a minimum at full load operations.

The control of HC and CO emissions is primarily a function of fuel injection and atomization and fuel-air mixing. Decreased HC and CO emissions are therefore accomplished by combustor and fuel injection modifications which promote better fuel atomization and fuel and air mixing. The chemical kinetics of combustion reactions show that HC compounds are consumed faster than CO, with the result that, as gas turbine efficiency is increased, any remaining non-equilibrium products of combustion will tend to exist mainly as CO. Therefore, reductions in HC and CO emissions can be obtained by controlling the residence time at temperature, as necessary, to provide combustion of HC in the primary zone of the combustor and combustion of CO in the primary and intermediate zones of the combustor.

The type of fuel burned can affect CO emissions. Tests by Westinghouse[1] indicate that higher CO emissions are produced by heavier fuels. This effect is reduced by proper design of the combustor to burn specific fuels.

7.4 SULFUR DIOXIDE EMISSIONS

SO_2 emissions from gas turbines are strictly a function of the fuel sulfur content, since virtually all fuel sulfur is converted to SO_2. The

only technique used at present to control SO_2 emissions from gas turbines is to burn low sulfur fuels. Stack gas scrubbing for SO_2 removal has not been applied to gas turbines primarily because of the large volumes of gas which have to be treated; EPA has expressed consensus with this conclusion.[3]

Practically all synfuels have specifications limiting the sulfur content to levels lower than the 0.8% maximum specified by EPA.

7.5 NITROGEN OXIDE EMISSIONS

Nitrogen oxides (essentially nitric oxide, NO) produced by combustion of fuels in stationary gas turbines are formed by the combination of nitrogen and oxygen in the combustion air ("thermal" NO_x) and by the combination of nitrogen in the fuel with oxygen from the combustion air ("organic" NO_x). Thermal nitric oxide formation rate is extremely sensitive to the flame temperature, increasing exponentially with increases in flame temperature. The exact mechanism of formation of organic NO_x is not known. Experiments by General Electric show that the actual amount of fuel bound nitrogen converted to NO_x decreases as the fuel nitrogen content increases, reaching a steady value of approximately 50% conversion at nitrogen contents of 0.3% or higher.

The following major control procedures can reduce NO_x emissions:

(1) Reduction of fuel bound nitrogen

(2) Injection of water or steam into

(3) Combustor modification

(4) Flue gas treatment

7.5.1 REDUCTION OF FUEL BOUND NITROGEN

Reduction of fuel bound nitrogen is achieved when the fuel is hydrotreated. Synfuels may exhibit nitrogen content ranging up to 1% or higher. Hydrotreating can lower the nitrogen content to 0.25% or less, thereby meeting the EPA standard. Additional advantages of this procedure are the upgrading of the fuel and the decreased biohazard (see below).

7.5.2 INJECTION OF STEAM OR WATER

The injection of steam or water into the combustor is a well established procedure achieving 70 to 90% reductions of thermal NO_x and more modest reductions of organic NO_x when a water/fuel ratio of 1.0 is used. Water injection reduces gas turbine efficiency by approximately 1%, while steam injection increases it by a similar amount.

7.5.3 COMBUSTOR MODIFICATION

Combustor modification techniques have been applied individually or in combination to reduce NO_x emissions. The following design modifications have been tested:

a. air staging and redistribution

b. fuel vaporization

c. fuel staging

d. two-stage combustion and off stoichiometric combustion

e. premixing of the air and fuel prior to introduction to the combustion chamber

f. variable combustor geometry

g. exhaust gas recirculation

h. catalytic combustion

i. external combustion in a larger combustion chamber(s) where the combustion conditions can be more easily controlled than in a conventional gas turbine combustor.

Many of these procedures are effective. The NASA-Lewis Research Center has sponsored a number of projects as part of its "Clean Combustor" program to demonstrate practical combustor technology for the reduction of pollutants in future generation aircraft turbines. Within this program, reductions of NO_x emissions up to 94% were obtained.

Pratt and Whitney performed for EPA during the period December, 1975 - November, 1979 an exploratory development program to identify, evaluate and demonstrate alternative combustor design concepts for significantly reducing the production of NO_x in stationary gas turbine engines.[4] Based on this program, the "rich burn-quick quench" concept, shown in Figure 7-17 of the Appendix volume, was selected for implementation into the design of a full-scale (25 megawatt engine size) gas turbine combustor. Preliminary test results showed that substantial reductions in NO_x from both nitrogenous and non-nitrogenous fuels could be obtained. The properties of the fuels and the NO_x emissions measured are presented in Table 7-1. As shown in the case of SRC-II middle distillate, acceptable NO_x emissions were generated by a fuel containing close to 1% nitrogen.

7.5.4 FLUE GAS TREATMENT

NO_x control can be achieved by post-combustion treatment of the flue gas with ammonia, with or without catalysts.[5]

Uncatalyzed reaction with ammonia is used in the Exxon Thermal $DeNO_x$ process, which has been applied for NO_x control in boilers and

furnaces.[6] In this process, ammonia is injected into the flue gas at a temperature range from 1600 to 1800°F; NO_x reductions of 70% are reported.

Hitachi (Japan) has developed catalysts resistant to SO_2 poisoning[7,8] which can reduce NO_x to nitrogen by reaction with ammonia in the presence of oxygen in a temperature range of 400 to 750°F. NO_x removal rates ranging up to 90 percent are claimed.

Flue gas treatment procedures have been applied mainly to conventional steam boilers rather than gas turbine operations, because the high velocity and high volume of turbine exhaust would require extremely large catalyst beds.

7.6 BIOHAZARDS

Carcinogenic compounds may form during direct liquefaction of coal and pyrolysis of oil shale; to a lesser degree, these compounds may also be present in petroleum resid. They typically have boiling points higher than 480°F, and consist mainly of polycyclic aromatic hydrocarbons and amines.

This synfuel biohazard affects mainly plant workers who come into direct contact with the fuels. Occasional exposure to the carcinogens is not sufficient for cancer development. Strict application of industrial hygiene practices is expected to avoid the development of any effects. Carcinogenic effects, even of a mild nature, such as skin cancer, have not ever appeared among the workers at the SRC-II Demonstration Plant at Tacoma, Washington, over many years of plant activity. This plant practices strict personnel protection.

A recent chemical and biological study of an SRC-II distillate blend[9] found that most of the mutagenic activity (related to carcinogenic activity), as revealed by the Ames test, could be attributed to primary aromatic amines. Hydrotreating of the fuel caused a significant reduction of the primary aromatic amines as well as of the polynuclear aromatic hydrocarbons, with concurrent reduction of mutagenic activity. Therefore, hydrotreating, which

is used to upgrade the fuel and reduce its nitrogen content, can also reduce its biohazard potential.

Use of non-hydrotreated high boiling synfuels or resid in gas turbines may lead to particulate emissions of unburned fuel on startup and shutdown. If further studies find these emissions hazardous, they could be avoided by burning distillate fuel on startup and shutdown, and switching to the heavier fuels when the turbine is operated at peak load and complete burning of the fuel is assured. This practice has already been followed with gas turbines burning heavy resid which has to be heated prior to use.

7.7 SECTION 7 LITERATURE CITED

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9. W. D. Felix et al. (Battelle Memorial Institute), "Chemical/Biological Characterization of SRC-II Product and Byproducts," presented at the Fifth Symposium on Environmental Aspects of Fuel Conversion Technology, St. Louis, Missouri, September 16-19, 1980.

Table 7-1 - Fuel Properties[a] and NO_x Emissions of Some Natural and Synthetic Fuels Used in the EPA "Rich Burn-Quick Quench" Program

Property	No. 2 (Typical)	SRC-II Middle Distillate	Indonesian/ Malaysian Resid	Shale Resid
Specific Gravity	0.84 (60°F)	0.97 (60°F)	0.87 (210°F)	0.82 (210°F)
Viscosity, centistokes	5.0 (60°F)	6.3 (60°F)	11.6 (210°F)	3.3 (210°F)
Surface Tension, dynes/cm	25.7 (60°F)	33.3 (60°F)	22.6[b] (210°F)	20.6[b] (210°F)
Heat of Combustion, (net) Btu/lbm	18,700	17,235	17,980	18,190
Pour Point, °F	< 5	<-45	61	90 (remains waxy)
Flash Point, °F	>130	>160	210	235
Ultimate Analysis				
Carbon %	87.0	85.77	86.53	86.71
Hydrogen %	12.8	9.20	11.93	12.76
Nitrogen %	< 0.02	0.95	0.24	0.46
Sulfur %	0.04-0.48	0.19	0.22	0.03
Ash %	< 0.003	0.001	0.036	0.009
Oxygen %	< 0.09	3.89	--	0.03
NO_x Emissions, ppm	40-45	90	75	65
Conradson Carbon, Residue %	< 0.30	0.03	3.98	0.19
Endpoint, °F (Atm Distillation)	640	541	NA	700

[a] Fuel properties are given at stand delivery temperatures to be maintained in test program.
[b] Estimate on basis of fuel specific gravity.